U.S. DEPARTMENT OF COMMERCE
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PB-282 090

Proceedings: Symposium on Flue Gas Desulfurization Held at Hollywood, Florida, November 1977. Volume I

Industrial Environmental Research Lab, Research Triangle Pk, N C

Mar 78

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/7-78-058a	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Proceedings: Symposium on Flue Gas Desulfurization--Hollywood, FL, November 1977 (Volume I)		5. REPORT DATE March 1978 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) Franklin A. Ayer, Compiler		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Research Triangle Institute P.O. Box 12194 Research Triangle Park, North Carolina 27709		10. PROGRAM ELEMENT NO. EHE624A 11. CONTRACT/GRANT NO. 68-02-2612, Task 38
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Industrial Environmental Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Proceedings; 11/8-11/77 14. SPONSORING AGENCY CODE EPA/600/13

15. SUPPLEMENTARY NOTES IERL-RTP project officer is Julian W. Jones, Mail Drop 61, 919/541-2489.

16. ABSTRACT The proceedings document presentations made during the symposium, which dealt with the status of flue gas desulfurization technology in the United States and abroad. Subjects considered included: regenerable, non-regenerable, and advanced processes; process costs; and by-product disposal, utilization, and marketing. The purpose of the symposium was to provide developers, vendors, users and those concerned with regulatory guidelines with a current review of progress made in applying processes for the reduction of sulfur dioxide emissions at the full- and semi-commercial scale.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pollution	Byproducts	Pollution Control	13B
Flue Gases	Disposal	Stationary Sources	21B
Sulfur Dioxide	Marketing		07B
Desulfurization			07A,07D
Regeneration			14B
Cost Analysis			

18. DISTRIBUTION STATEMENT Unlimited	19. SECURITY CLASS (This Report) Unclassified	21. NO. OF PAGES 440
	20. SECURITY CLASS (This page) Unclassified	22. PRICE A19- A01

EPA Form 2220-1 (9-73)

☆U.S. GOVERNMENT PRINTING OFFICE: 1978 -740 -261/ 337 REGION NO. 4

PB 282 090

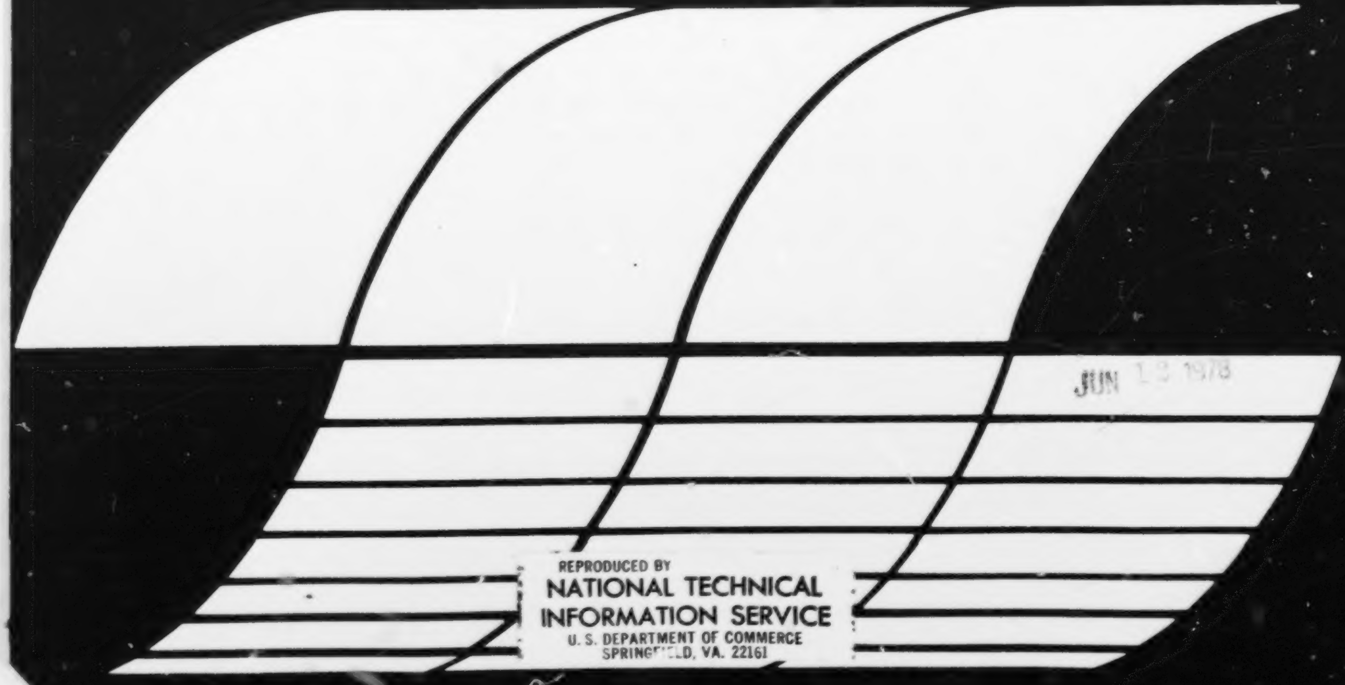
EPA
U.S. Environmental Protection Agency
Office of Research and Development

Industrial Environmental Research
Laboratory
Research Triangle Park, North Carolina 27711

EPA-600/7-78-058a
March 1978

PROCEEDINGS: SYMPOSIUM ON FLUE GAS DESULFURIZATION— Hollywood, FL, November 1977 (Volume I)

Interagency
Energy-Environment
Research and Development
Program Report



JUN 1 2 1978

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EPA-600/7-78-058a
March 1978

PROCEEDINGS: SYMPOSIUM ON FLUE GAS DESULFURIZATION— Hollywood, FL, November 1977

(Volume I)

Franklin A. Ayer, Compiler

Research Triangle Institute
P. O. Box 12194
Research Triangle Park, N. C. 27709

Contract No. 68-02-2612
Task 38
Program Element No. EHE624A

EPA Project Officer: Julian W. Jones

Industrial Environmental Research Laboratory
Office of Energy, Minerals and Industry
Research Triangle Park, N.C. 27711

Prepared for

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Research and Development
Washington, D.C. 20460

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PREFACE

More than half of all "man-made" sulfur dioxide (SO_2) is emitted by electric power plants, and the use of sulfur-containing fossil fuels, especially coal, to generate electricity is expected to increase dramatically in the next 10 years. To avoid the adverse environmental effects of this increase in fossil fuel combustion, the development and commercial application of SO_2 control technologies is one of the most important concerns of the U.S. Environmental Protection Agency (EPA). Flue gas desulfurization (FGD) is the most promising technique for control of SO_2 that will be available for widespread application to fossil fuel-fired electric power plants for at least the next 8 to 10 years.

The Industrial Environmental Research Laboratory - Research Triangle Park (IERL-RTP) of EPA's Office of Research and Development sponsors symposia for the transfer of information regarding FGD research, development and application activities with the objective of further accelerating the development and commercialization of this technology. These symposia provide an opportunity for users and developers to discuss their experiences and the status of development and application of FGD technology.

The November 1977 symposium addressed full-scale FGD process applications in the United States, Japan, and West Germany, as well as laboratory, pilot, and prototype research and development efforts. The symposium also provided an opportunity for the announcement of data and results which were previously unreported or not widely publicized. The economics of FGD and the disposal, utilization, and marketing of FGD system byproducts were also discussed. The symposium papers were presented by a cross section of those concerned with FGD including users, government and private developers, and suppliers. The electric utility industry—the principal user of FGD—participated extensively in the symposium program. More than 800 people attended the symposium.

The General Chairman of the November 1977 Symposium on Flue Gas Desulfurization was Michael A. Maxwell, Chief, Emissions/Effluent Technology Branch, IERL-RTP. The Vice Chairman was Julian W. Jones, a Chemical Engineer in the Emissions/Effluent Technology Branch, IERL-RTP.

These Proceedings are comprised of copies of the participating authors' papers as received. As supplies permit, copies of the Proceedings are available free of charge and may be obtained by contacting IERL-RTP's Technical Information Coordinator, Environmental Protection Agency, Research Triangle Park, North Carolina 27711.

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REMARKS

Stephen J. Gage

Acting Assistant Administrator for
Research and Development
U.S. Environmental Protection Agency
Washington, D.C.

REMARKS BY

STEPHEN J. GAGE

Before the Fourth Symposium on Flue Gas Desulfurization

I am delighted to return again to the Symposium on Flue Gas Desulfurization; this is my fourth appearance. It struck me that not only is this my fourth appearance, but also my fourth time in a different role. Oh well, some people just can't hold a steady job.

Each time I've come to the Symposium I've had a special interest in the subject and a special regard and affection for the people who have put the Symposium together year after year. John Burchard, Bill Plyler, Mike Maxwell, Dick Stern, Julian Jones, Bob Borgwardt, Wade Ponder, Norm Kaplan and the rest of the good people from the Industrial Environmental Research Laboratory and Frank Princiotta of my Headquarters Office deserve lots of credit for providing one of the most effective means of technology transfer that EPA has undertaken. It's good to see some old EPA alumni here such as Kurt Yeager and Gary Rochelle. And lots of other old friends and collaborators--Bill Elder and Professor Jumpei Ando of Japan, for example.

But there are many new faces, too. While Dr. Ando has been bringing status reports from Japan for several years, we have some new participants from the Federal Republic of Germany--Drs. Rolf Holighaus and Michael Esche--to whom I want to extend my personal welcome and thanks for hosting us in

Germany during the past years. More about that later. I also want to express my appreciation to those of you from utilities and industries bringing to the Symposium the most recent developments from operating scrubber facilities. I wish you well in this important conference.

Before I get on with my assigned task this morning--that of introducing the Keynote Speaker--I would like to share with you some recent funding developments in our pollution control program for coal-fired boilers. I recall that I had a similar opportunity just about three years ago when I described to the Symposium the establishment of the EPA-coordinated Federal Interagency Energy/Environmental Program.

As all of you know, one of the major facets of the National Energy Plan announced by President Carter this past spring is a dramatic increase in the use of our vast supplies of coal. This increase in coal use, combined with major conservation efforts and development of non-polluting energy sources such as solar, is part of a comprehensive plan to alleviate the current energy crisis, a crisis revealed in the fact that half of the oil we currently consume is being supplied by imports. To meet the goals of the National Energy Plan, a major expansion in coal production is called for--from less than 100 million tons in 1976 to an annual production of over one billion tons by 1985. In addition, the plan calls for conversions of existing oil- and natural gas-consuming plants to coal, for generating both electric power and industrial heat.

In view of past conversions from coal to oil and gas in order to meet the standards for cleaner air, it is obvious that the shift back to coal and the increasing demand for energy will markedly expand the potential for environmental degradation. Concern over these potential problems is heightened by the concentration of basic manufacturing and population centers in relatively small areas of the country such as the industrial northeast and midwest. Many utility and industrial boilers are in these areas, resulting in high atmospheric concentrations of combustion pollutants, often exceeding national ambient air quality standards.

In addressing these potential problems, the President stated in his Energy Plan "...that all new facilities, including those that burn low sulfur coal, should be required to use the best available control technology."[1] But I'll leave the discussion of these developments to our distinguished Keynoter.

I would like to point out that the National Energy Plan also calls for "A comprehensive coal research and development program...(which)...should focus on meeting environmental requirements more effectively and economically..." The Plan goes on to say that "increased research will be devoted to control the fine particulate and sulfur oxide emissions with coal burning..."[2]

Responding to the President, EPA submitted a supplemental appropriation request for an accelerated program for coal polution control technology development, demonstration, and information dissemination. The $35 Million request was approved by the Office of Management and Budget, and submitted to the Congress. The House and Senate Conference Committee recently approved a compromise appropriation of $34.7 Million, while the appropriation is currently tangled up in the Presidential veto of the Clinch River Breeder appropriation, we expect the approval by both houses and signature by the President in the next few weeks. The final package approved by the Conference Committee differs slightly from those submitted by EPA in that a $4 Million program to characterize atmospheric nitrates was deleted and a $3.7 Million program to accelerate research on coal cleaning was added.

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[1] The National Energy Plan, page 67.

[2] The National Energy Plan, page 68.

The bulk of the supplemental appropriation--$31 Million--is designed to ensure the availability of effective air pollution control alternatives for combustion sources and to reduce the cost of using these options. Specifically, in fiscal year 1978, an additional $12 Million will be made available for nitrogen oxides control, $11 Million for fine particulate control, and $8 Million for sulfur oxides control.

I would now like to describe briefly what is contained in each of these three program areas. In the area of nitrogen oxides control, we will emphasize:

--demonstrating the reliability of currently available combustion modifications in full-scale boilers, with special concern for the boiler tube corrosion;

--broadening the application of combustion modifications technqiues to industrial boilers; and

--developing and demonstrating improved burner designs which promise reductions of up to two-thirds of the NO_x emitted from large boilers today.

We believe that this is a vital area which must yield to hard-nosed research efforts in order to check the rapidly rising emissions of NO_x nation-wide. Break-throughs in this area are extremely important because of the slow-down in control of NO_x from automobiles and the implications of NO_x in health effects, acid rains with high nitric acid content, and photochemical smog in urban areas.

The fine particulate control efforts will advance the application of fabric filters, electrostatic precipitators, and wet scrubbers to the control of submicron particulate emissions from coal-fired power plants. The use of fabric filters to collect fine particles from coal plants is growing and there is evidence that even gaseous sulfur oxides may be captured through the injection of highly reactive absorbency into the baghouse itself. Conditioning techniques--both chemical and electrical--may improve the ability of ESP's to collect fine high resistivity fly ash usually associated with lower sulfur natural or cleaned fuels. Finally, the ability of wet scrubbers to more effectively capture fine particles through the use of condensation phenomena appears likely.

The supplemental appropriation for flue gas desulfurization will be primarily focused on helping solve these remaining problems which could be barriers to the application of current-generation FGD technologies, in other words, a shot in the arm for those efforts which EPA, the Tennessee Valley Authority, the Electric Power Research Institute, and many of you are now carrying one. We will incidentally be working closely with both TVA and EPRI in all of these expanded efforts. We'll also be cooperating with the Federal Republic of Germany in most of these areas under the US-FRG Environmental Agreement; these joint activities have been given a real boost on the German side as well as by a major financial commitment by the German government to reduce pollution from coal-fired plants.

We'll be concentrating on those technologies such as lime/limestone scrubbing which have already had significant market penetration as well as those which are now moving into commercialization such as Double Alkali ad Wellman/Lord. Through these cooperative efforts, we're relatively confident that most of the nagging problems of such scrubbers can be overcome and that these FGD systems can be installed and operated by utilities and industries with little risk.

That, in a nutshell, is our ambitious expansion of the air pollution control program for coal-fired boilers. I am hopeful that our objectives can be achieved in the next few years. Through such a cooperative Government/private industry approach, we will certainly be maximizing our chances of success.

It remains for me today to introduce a distinguished colleague of mine as your Keynote Speaker.

David Hawkins, who is even younger than I, has already had a very significant impact on the use of the Clean Air Act to reduce air pollution in the United States. As an attorney with the Natural Resources Defense Council from 1971-1977, David became one of the leading experts in the country on control of air pollution from power plants, steel works, and smelters, on attainment of air quality standards, on transportation and indirect source controls, and on new-source reviews. He was frequently involved in EPA's regulatory proceedings as well as those of other Federal agencies and often testified before Congressional committees.

I've the opportunity to work with him for the past six months. I've found him intelligent, articulate, fair, committed--in short, a prince of a guy to work with. I'm pleased to work with him. And I'll think you'll appreciate him even more once you've heard him. David Hawkins, your Keynote Speaker--

KEYNOTE ADDRESS: THE CLEAN AIR ACT AMENDMENTS OF 1977 - NEW DIMENSIONS IN AIR QUALITY MANAGEMENT

David G. Hawkins

Assistant Administrator
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U.S. Environmental Protection Agency
Washington, D.C.

The Clean Air Act Amendments of 1977 - New Dimensions in Air Quality Management

by

David G. Hawkins

I'm very glad to be here today. You know, this is my first speech to a major group since taking my new job at EPA, so I feel privileged that you are the audience. The people assembled here know first hand of the practical difficulties of reducing emissions. I have a feeling that this detailed knowledge gives you the opportunity to make a big difference in just how clean our air is and will be, perhaps, a bigger difference than us regulation writers in Washington. The status of FGD today is a tremendous personal achievement by each of you who has been involved. You should feel proud.

The list of papers on the agenda is impressive. I wish I had time to hear or read all of them. I'm sure the results of these papers will reach us though and will make a big difference in my work.

This morning I would like to highlight a number of the important changes in our basic federal air pollution law made by the Clean Air Act Amendments of 1977.

The recent Amendments provide a strong mandate, clearly indicating that Congress does not intend that the nation's energy problem or economic problems be allowed to compromise environmental quality. Some people have referred to the new legislation as a "mid-course correction." I believe that "mid-course reaffirmation" is more appropriate. The Amendments strongly reaffirm our goals of health standards attainment and the prevention of significant air quality deterioration. They renew and strengthen our existing programs, as well as add new programs.

The amended Act requires EPA to examine the need to control pollutants other than those covered by national ambient air quality standards and to review and modify, if necessary, existing regulatory initiatives. It emphasizes the development of adequate air pollution control programs at the state level, Federal assumption of enforcement responsibilities if necessary, and where inaction continues, the Act provides for strong sanctions.

The Clean Air Act Amendments of 1977 provide a major incentive to resolve potential conflicts between environmental goals and energy, economics, and employment concerns. The focus is one of early identification of obstacles and intensified efforts to solve these problems in the planning phase of new facilities. Now I would like to discuss a few of the more important provisions of the Amendments.

One of the most significant concerns addressed by the Congress is the problem of attainment in areas where the National Ambient Air Quality Standards continue to be violated. Congress took a firm position on this issue. Each State that includes a non-attainment area must submit for EPA approval, a revised plan which provides for attainment of the standards as soon as practicable, but not later than December 1982. However, Congress recognized unique problems associated with photochemical oxidants and carbon monoxide, and authorized the Administrator to grant some extensions to December 1987. In many cases, the new State plan requirements will require the application of the most stringent controls available and strict enforcement of these requirements for existing sources. The plan will have to provide sufficient reductions in emissions to allow room for new growth, otherwise construction permits for new sources can be granted only when more than offsetting emission reductions are secured on a case-by-case basis prior to the facility startup date. This is a challenge--a challenge we are committed to meet; states will be asked to do what is possible, but they will be asked to do their job well. Public health is at stake and we don't intend to be timid in our efforts to achieve it.

There are currently over 1000 major emitting facilities that even now are not in compliance with requirements under the Clean Air Act. The Amendments establish new enforcement tools which can provide an expeditious but equitable cleanup schedule. These requirements govern any enforcement orders issued by either State or Federal authorities which extend the time for compliance. Such orders must contain final compliance dates that are an expeditious as practicable, but in no event later than 3 years from the date that attainment of the national ambient air quality standards would otherwise have been required. Should a source wish to comply by use of innovative technology, however, it may be granted up to 5 years to achieve final compliance. A source would not be eligible for a delayed compliance order if the delay is requested solely for convenience or for economic advantage.

To ensure that these delayed compliance orders are not used as a mechanism for gaining economic or other competitive advantage, the issuance of such order will serve as a point of reference for purposes of establishing non-compliance penalties should the affected source not comply by the applicable date. Using the variables set forth in the Amendments, the non-compliance penalty would be calculated on the basis of the costs which a non-complying source avoids by delayed compliance. Specifically, the calculation of the penalty must consider the capital costs of compliance and debt service over a normal amortization period not to exceed ten years, operation and maintenance costs foregone, and any other additional economic value of delay. In essence, the penalty is intended to reflect any financial savings realized by a firm as a result of non-compliance with the law, and should eliminate any economic advantage that delayed compliance will confer on a noncomplying firm relative to a firm that complies in a timely manner.

With respect to the means of compliance available to a source for purposes of conforming with the air quality goals, Congress again took a firm stand. The Congress reaffirmed its commitment to the mandate of the 1970 Amendments that atmospheric loading through dispersion techniques (either spatial or temporal) is not an acceptable means of meeting air quality goals. The use of tall stacks or intermittent control systems for dispersion of air pollutants will not be allowed for purposes of determining compliance with national ambient air quality standards or PSD increments.

With regard to the national energy problem confronting us today, the Congress has indicated its belief that the pursuit of less reliance on foreign energy supplies be compatible with a clean environment. The Congress acknowledged the need to expand our use of domestic coal to meet national energy needs but specified that this be done without reducing our commitment to improved environmental quality. The amendments therefore do not provide for relaxation of any air quality standards but provide mechanisms to permit the continued substitution of domestic coal for oil and gas. The coal conversion provisions basically the same authorities and responsibilities that were established for EPA under the Energy Supply and Environmental Coordination Act of 1974.

However, one section of the new amendments now deals further with the issues of economic disruption and unemployment. This section provides that sources burning coal or those ordered to convert to coal can be directed to use locally available supplies rather than importing their fuel from other regions. Recognizing that in some instances this may involve the burning of high sulfur coal, the provision also requires the use of pollution control technology to keep emissions at the lowest levels. The emphasis on continuous emission reduction and best available control technology in the new Act, which Steve Gage referred to earlier, is of particular importance to this group.

Section III of the Clean Air Act, dealing with standards of performance for new stationary sources, was amended to require an emission standard '...requiring the achievement of a percentage reduction in emissions from such category of sources from the emissions which would have resulted from the use of fuels which are not subject to treatment prior to combustion." This percentage of reduction is further defined as the "...application of the best system of continuous emission reduction which (taking into consideration the cost of achieving such emission reduction, and any non-air quality health and environmental impact and energy requirements) the Administrator determines has been adequately demonstrated for that category of sources."

EPA has begun to implement Congress's mandate with respect to fossil-fuel fired steam generators. The Agency expects to propose a revision to the new source performance standards for steam generators in early 1978 with promulgation following in August.

Preliminary work indicates that the standard for sulfur dioxide will contain three parts:

(1) a percentage reduction, as required by the Act,
(2) maximum allowable emissions (a "ceiling"), and
(3) a maximum control level (an emission limit achievable by untreated combustion of a very low sulfur fuel)

The current staff draft calls for 90 percent control with 90 percent availability; however, these issues are not yet resolved.

The second part of the standard places a ceiling on allowable emissions. One alternative identified by the staff is to establish this limit at 1.2 pounds of SO_2 per million Btu heat input, the current new source performance standard. This means that certain very high sulfur content coals could not be used with even a high efficiency FGD system. It is possible that such fuels could be used if they were sufficiently precleaned before firing.

The third part of the standard will establish a maximum control level requirement. The staff draft identifies 0.2 pounds SO_2 per million Btu heat input as one possible floor. This is based on the recognition of the low SO_2 emission characteristic of fuels with a very low sulfur content. Such fuels could be used in combination with other fuels to reduce overall SO_2 control requirements. The emission limit is based on a 90 percent reduction in the potential emissions of a high quality Western coal. Fuels with a very low sulfur content (natural gas, wood residue, some municipal waste, etc.) can achieve this emission level without additional SO_2 control. Other marginal fuels with emission rates only slightly in excess of this level (low sulfur fuel oil, some municipal waste, etc.) would require some SO_2 emission control or could be fired in combination with a very low sulfur fuel with no SO_2 control.

As you can realize, the time table for proposal and promulgation of the standard is extremely tight. A great deal of effort has been expended, and significant resources are committed to analyze issues relative to the impact of such standards on the fuel industry and on the availability of control technology. The research and development community, public and private, is an indispensable partner in these efforts, and I am grateful for their contribution.

Continuous emission reduction and best available control technology requirements are also relevant to another portion of the Clean Air Amendments of 1977: Part C - Prevention of Significant Deterioration of Air Quality (PSD). I believe it is useful to summarize the purposes of this part as stated in the Act:

"(1) to protect public health and welfare from any actual or Potential adverse effect...from exposures to pollutants in other media, which pollutants originate as emissions to the ambient air,

(2) to preserve, protect, and enhance the air quality in areas of special national (or regional) scenic, natural, recreational, or historic value,

(3) to ensure that economic growth will occur in a manner consistent with the preservation of existing clean air resources,

(4) to assure that emissions from any source in any State will not interfere with the prevention of significant deterioration of air quality in any other State, and

(5) to assure that any decision to permit increased air pollution in any area to which this section applies is made only after careful evaluation of all the consequences of such a decision and after adequate procedural opportunities for informal public participation in the decision-making process."

These are pretty sweeping. The new provisions for the prevention of significant deterioration modify and substantially expand EPA's former regulations dealing with PSD established under the 1970 version of the Clean Air Act. Some of the most significant changes to EPA's former PSD program are as follows:

The Act automatically classifies certain areas of the country as Class I and therefore subject to the most stringent constraints on air quality deterioration. These areas include all international parks, all national wilderness areas and memorial parts which exceed 5000 acres in size, and all national parks which exceed 6000 acres in size; about 150 areas in all.

The Act sets forth new, more restrictive, ambient air quality increments for particulate matter and sulfur dioxide in Class II and Class III areas. Also newly effective is the requirement that each national ambient air quality standard shall act as an overriding ceiling to any otherwise allowable increment.

The enforcement tool which provides the basis for the PSD program is the issuance of preconstruction permits which must be obtained by all major new sources of air pollution. Part 165 of the Act requires that the proposed facility be subject to the best available control technology for each pollutant subject to regulation under the Act, which is emitted from the facility. As you may have heard there is some dispute about when this is to be implemented. We will make this effective in March 1978.

One particular feature of the Amendments dealing with the prevention of significant deterioration is the establishment of new national goal: the prevention of any future (and the remedying of any existing) impairment of visibility in mandatory Class I areas where impairment results from man-made air pollution. The Amendments require the Administrator to promulgate regulations to assure reasonable progress toward meeting the national goals. The regulations will include guidelines for the states and require each applicable state implementation plan to contain emission limits, compliance schedules, a long term strategy and other measures as may be necessary. One provision of the Act requires certain existing major stationary sources which emit any air pollutant which may reasonably be anticipated to contribute to impairment of visibility, to install the best available retrofit technology for controlling emissions. Congress specifically identified fossil-fuel fired generating powerplants having a total generating capacity in excess of 750 megawatts as subject to the best available retrofit requirement.

As you can see from this brief discussion of key provisions of the 1977 Clean Air Act Amendments, the Congress is even more concerned today with environmental protection than in 1970. Clean air is not an aesthetic luxury; it is a public health necessity. Many regions of the country have not yet met the health-based primary ambient air quality standards. These amendments define more sharply the regulatory tools needed to attain and maintain the standards.

As you discuss and deliberate issues relating to control technology in these next three days, you will be moving us one step closer to our goals. I wish you success with your symposium.

Thank you.

OVERVIEW SESSION

Session Chairman

Michael A. Maxwell
Chief, Emissions/Effluent Technology Branch
Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, North Carolina

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STATUS OF FLUE GAS DESULFURIZATION SYSTEMS IN THE UNITED STATES

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ABSTRACT

PEDCo Environmental, under contract to the Industrial Environmental Research Laboratory/RTP and the Division of Stationary Source Enforcement of the U.S. Environmental Protection Agency, has been monitoring the status of flue gas desulfurization (FGD) systems since 1974. The information needed for this program is obtained by visiting operational FGD systems and through periodic contacts with representatives of utility companies, FGD system suppliers, design engineering firms, and regulatory agencies. This paper summarizes the results of this program, including:

(1) The current status of FGD applications in the utility sector, identifying the number of operational systems, as well as those planned or under construction as a function of system (process) type;
(2) The trend in overall FGD system performance; and
(3) Areas of technological advancement including process design changes in the first, second, and third generation FGD systems.

SECTION 1

INTRODUCTION

PEDCo Environmental, under contract to the EPA, has closely monitored the growth and use of FGD technology by utilities in the United States. This program involves FGD technology evaluations on both an overview and site specific basis. Site-specific analyses are based upon visits to operating FGD systems where process design information as well as data related to operational problems and their solutions, and capital and operating costs are obtained. A series of reports on major installations has been issued and reports on other systems are in preparation.

The most visible part of this project, however, is related to the periodic contacts made to prepare the bimonthly status reports. These reports summarize data on the number of systems, and their capacity, in operation, under construction or planned, as well as describe the performance of operating systems during the reporting period. This information is contributed by the utility industry, system suppliers, and other representatives, to further the timely transfer of information in this key technology area. Information provided by representatives of operating systems is reported essentially as obtained with little attempt made to further analyze or interpret the data. Information provided by system suppliers and other sources is confirmed with the appropriate utility prior to publication.

The following sections of the paper address some of the highlights of this survey program.

SECTION 2

NUMBER OF SYSTEMS - OPERATING AND PLANNED

The total number of installations, both active and inactive, and their equivalent electrical MW capacity as of August 1977 is summarized in Table 2-1.

Table 2-1. NUMBER AND CAPACITY OF U.S. UTILITY FGD SYSTEMS

Status	Number of units	Capacity, MW
Active		
Operational	29	8,914
Under construction	28	11,810
Planned		
Contract awarded	23	11,880
Letter of intent signed	5	1,892
Requesting/evaluating bids	5	2,825
Considering FGD (preliminary design stage)	35	16,031
Inactive	16	1,488
Total	141	54,840

As the table shows, there were 141 systems with an equivalent electrical capacity of 54,840 MW, including both active and inactive systems. Of the active systems, 29 were operational (8,914 MW); 28 were under construction (11,810 MW); and 68 systems were planned (32,628 MW). An additional 16 installations (8592 MW), which are considering the use of FGD as well as other control strategies such as complying sulfur content coal, are not included. In addition, a number of plants that are definitely planning to use FGD systems are omitted because such information is not ready for public release. Approximately 12 to 15 systems and 6000 MW are in this category. To date, 16 systems (1,488 MW) have been shut down for various reasons. Several of these systems were considered demonstration systems whereas others were based upon first generation technology.

Growth Trends

Figure 2-1 illustrates both the number and equivalent capacity of systems as a function of year of start-up. The "number of systems" requires clarification. A system is defined on the basis of inlet ducting configuration. A module or several modules which are ducted to one or more boilers is defined as a system. Thus a single FGD module which treats flue gas from only one boiler is considered a system just as four boilers which have a common inlet ducting system to say eight modules would also be considered as one system. On the other hand, a plant with five boilers that are ducted to five distinct modules, or groups of scrubber modules, with no common ducting between these five modules, would be considered as five systems.

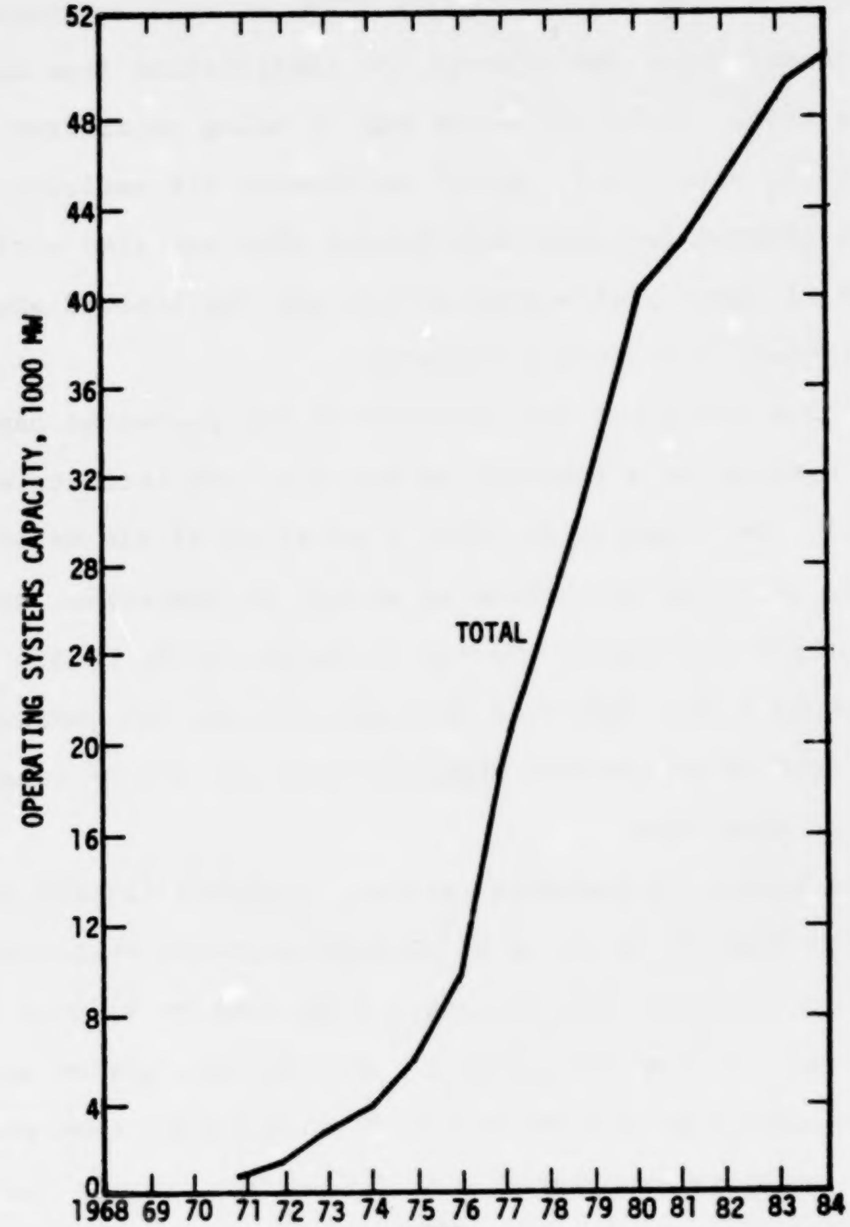

Figure 2-1. FGD operating capacity through 1984.

Figure 2-1 is based upon all the FGD systems installed and operated from 1968 to August 1977, as well as those under construction and planned for installation from August 1977 to 1984. Units for which FGD is being considered in conjunction with other control techniques are excluded. Systems planned for operation beyond 1984 are also excluded, because of their preliminary nature and the limited amount of publically available information.

Figure 2-2 shows the increase in the projected capacity of FGD systems as a function of the year the estimate was prepared. For example in 1974, a total of 37,836 MW of capacity could be identified as either in operation, under construction or planned whereas in August 1977, 53,352 MW are accounted for, and this does not include the approximately 6000 MW of planned capacity that can not be identified at this time.

The number of operating systems reported in 1974 has increased from 19 to 29, a 53 percent increase while the equivalent capacity has increased from 3291 MW to 8914 MW, an increase of over 170 percent. The average system size has increased from 173 MW to 307 MW in the same time period, and the capacity associated with full-scale systems has increased from 2360 MW to 8363 MW. Full scale systems are

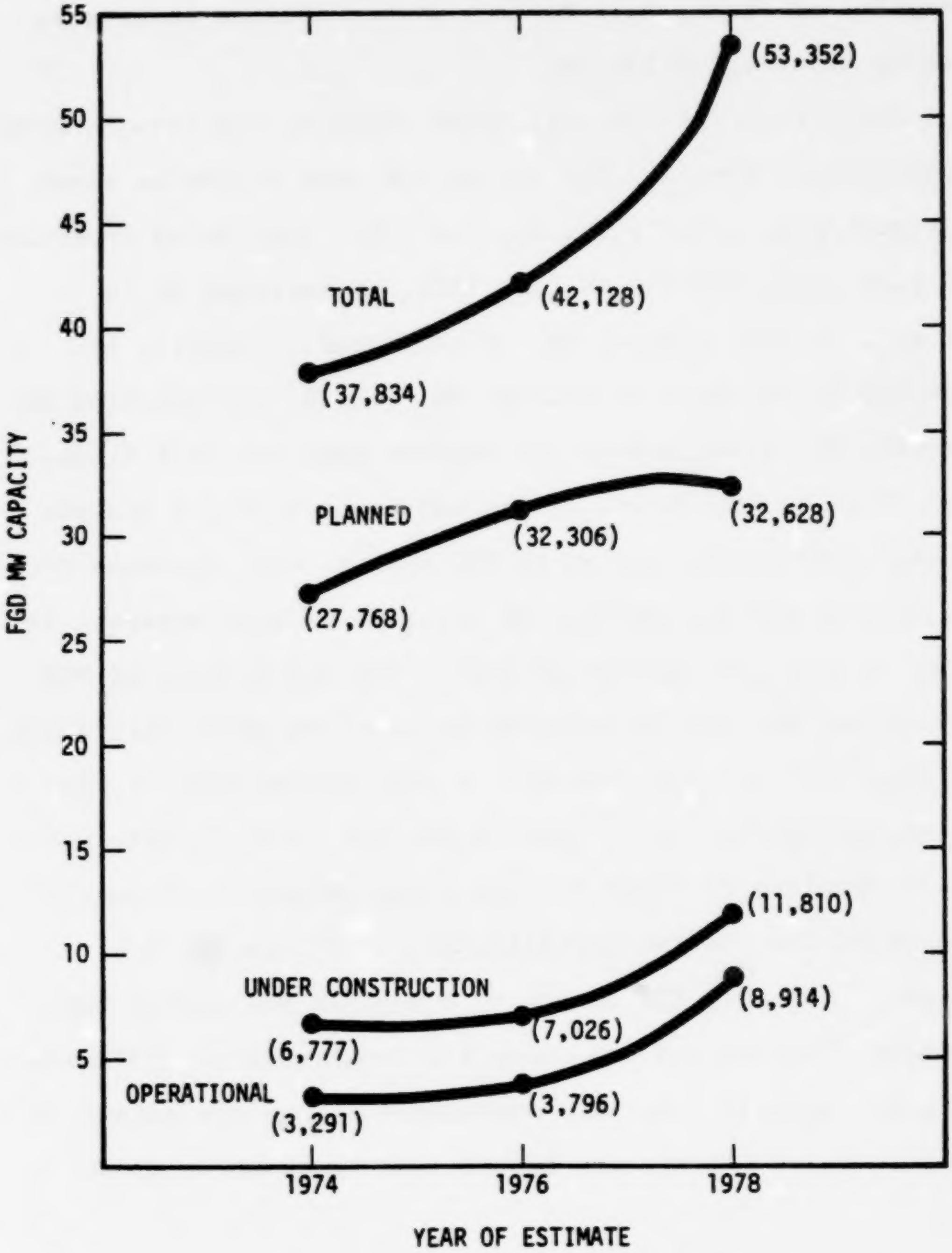

Figure 2-2. Increase in FGD utilization as a function of the year th. stimate was made.

defined as those systems that are available for commercial operation on fossil fuel boilers with a minimum power generating capacity of 100 MW.

Projections of new coal fired capacity are fraught with uncertainty. However, one projection used by EPA is shown in Figure 2-3. This shows that in 1975, coal fired capacity was just under 200,000 MW. By 1985, an increase of 70 percent, to over 330,000 MW, is projected. Capacity is expected to increase to 430,000 MW by 1990, representing an increase of approximately 120 percent over the 1975 figure. Also shown on Figure 2-3 is the application of FGD systems through 1984. Application of FGD systems will increase from the approximately 3 percent of total coal fired capacity in 1975, to about 15 percent in 1980. The utilization of FGD systems beyond 1980 is expected to increase above that shown on Figure 2-3 for two reasons. First, projections of FGD system application in the period between 1980 and 1984 are not as complete as those for new power generating capacity because of the shorter installation lead times for FGD systems. Secondly the impact of a revised New Source Performance Standard and the Clean Air Amendments of 1977 have not been factored into this assessment. Thus the extent of application of FGD systems beyond 1980 is probably under-

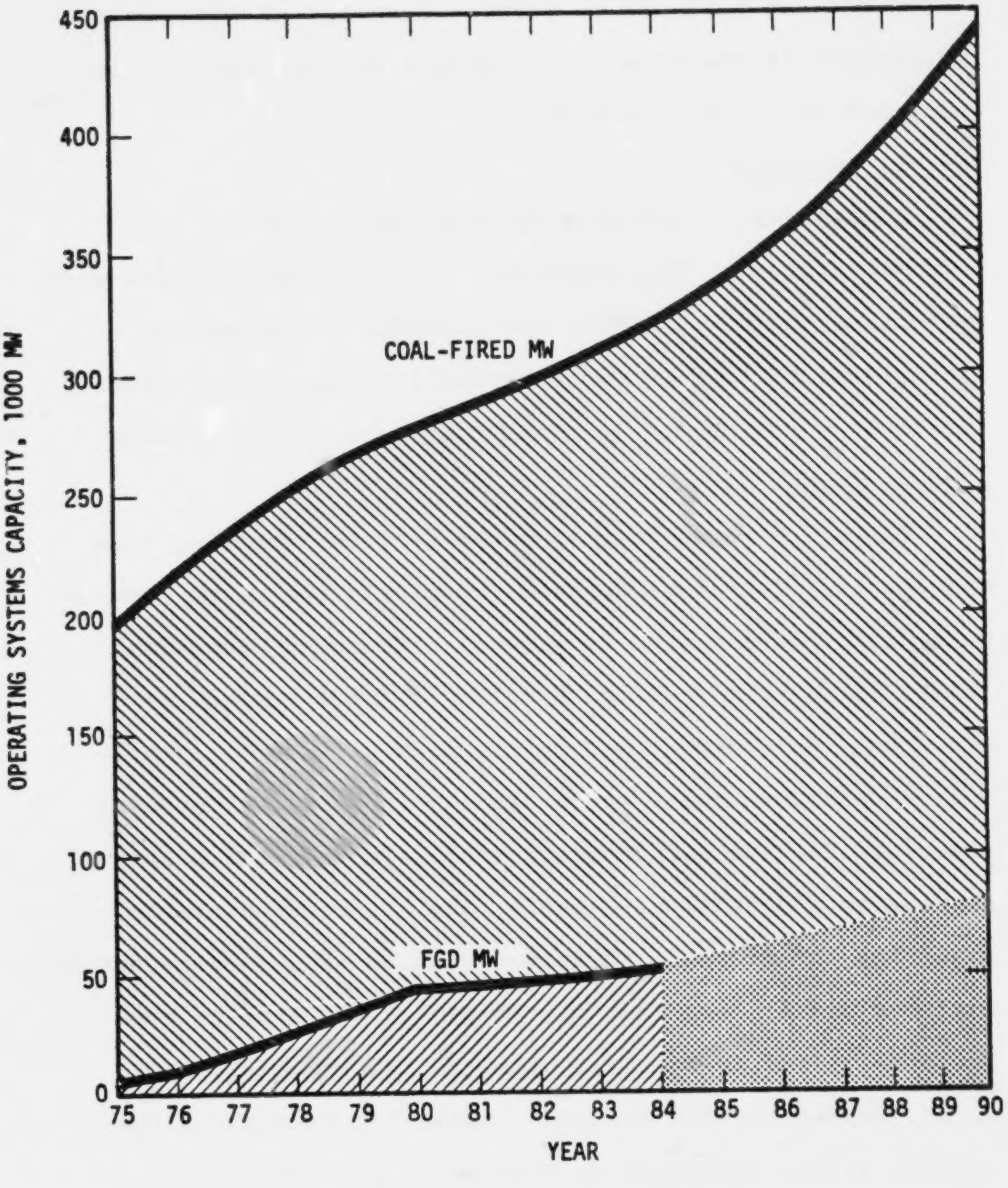

Figure 2-3. Increase in coal-fired capacity from 1975 to 1990 and corresponding increase in FGD capacity from 1975 to 1984.

stated and the tail-off in percent increase of FGD system application is due more to the method of capacity projection, per se, than a real trend.

New vs. Retrofit

Figure 2-4 illustrates the application of new vs. retrofit systems. The majority of initial applications were retrofit systems. In 1975, for example, 60 percent of operational FGD systems were retrofit applications whereas in 1980, approximately 70 percent will be on new systems.

Process Types

There are three primary methods of categorizing flue gas desulfurization processes: physical mechanism, chemical mechanism, and end product system. "Physical mechanism" refers to the phase in which sulfur dioxide removal is effected, i.e., either wet or dry phase. "Chemical mechanism" refers to the reagent used. "End product" systems are classified as either regenerable, recovery of sulfur dioxide in a usable, marketable form, or nonregenerable, requiring the disposal of sulfur dioxide as a nonrecoverable waste material. The vast majority of FGD operating experience gained to date, is with the calcium-based, wet phase, nonregenerable systems. Table 2-2 summarizes the systems either operating or planned by process type.

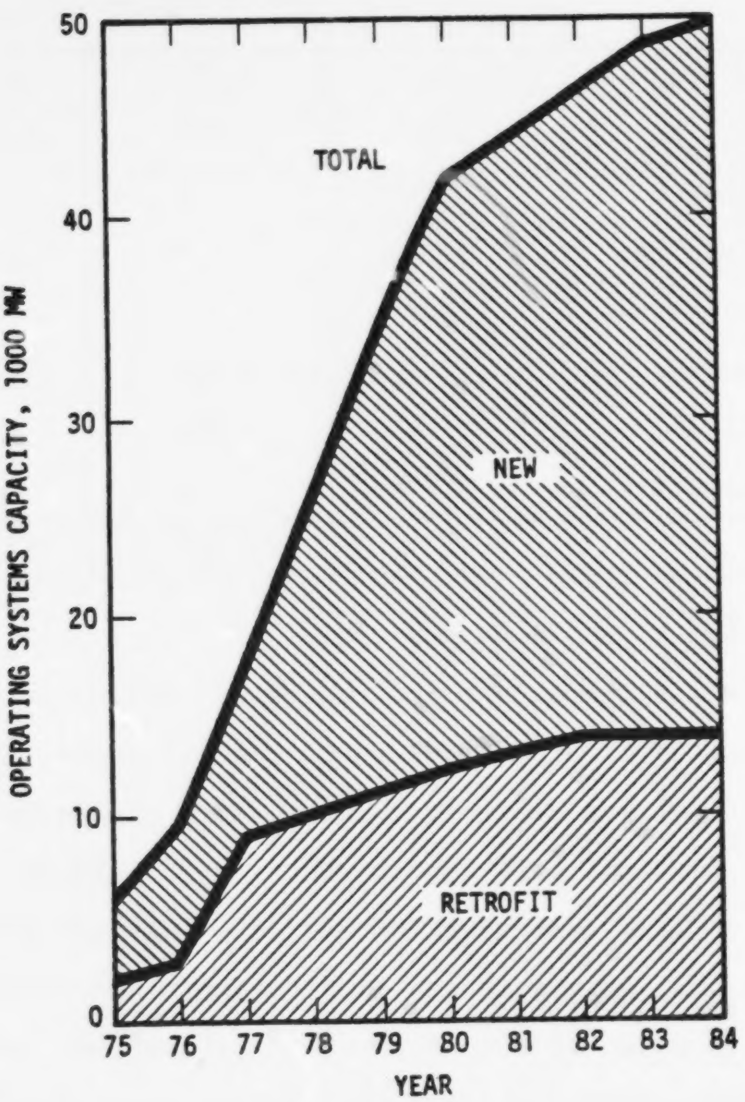

Figure 2-4. FGD operating capacity for new and retrofit installations through 1984.

Table 2-2. CHEMICAL PROCESS TYPES OF THE ACTIVE FGD SYSTEM CATEGORIES

Process	FGD capacity, MW			
	Operational	Construction	Planning	Total
Limestone	4,047	6,470	7,486	18,003
Lime	4,237	3,773	5,942	13,952
Lime/limestone	20	0	0	20
Sodium carbonate	375	0	634	1,009
Magnesium oxide	120	0	726	846
Wellman Lord	115	715	180	1,010
Double alkali	0	852	250	1,102
Aqueous carbonate	0	0	100	100
Total	8,914	11,810	15,318	36,042

Emission Limiting Standards

Table 2-3 summarizes the systems according to regulatory standards they must meet. Of the 125 active systems, 57 (23,930 MW) are designed to meet state standards more stringent than the NSPS requirement; 44 (22,728 MW) are designed to meet Federal NSPS; and 21 (5,819 MW) are designed to meet regulations less stringent than Federal NSPS. It is interesting to note that of the 29 active, operational systems, MW), more than half the systems (approximately two-thirds of the equivalent MW capacity) are meeting standards more stringent than current Federal NSPS.

Table 2-3. NUMBER AND CAPACITY OF ACTIVE FGD SYSTEMS FOR REGULATORY CLASSIFICATION CATEGORIES

Regulatory classification	Systems	Capacity, MW
Federal NSPS	44	22,728
More stringent than Federal NSPS	59	23,930
Less stringent than Federal NSPS	21	5,819
Undetermined	3	875
Total	125	53,352

High vs. Low Sulfur Coal Application

The design and operation of FGD systems for high- and low-sulfur coal is another area of interest, especially with regard to the viability of the systems on high sulfur coal applications. Because of the ambiguities inherent in the terms "high" and "low" sulfur coal, we have defined them for purposes of this paper as follows: low sulfur coal is any coal that when combusted will emit equal to or less than 1.2 lb SO_2/MM Btu, the current Federal New Source Performance Standard, and a high sulfur coal as any coal that will result in a higher emission value. Using these definitions, the following observations hold:

- Among the operating systems, approximately 85 percent of the equivalent electrical MW capacity is on high sulfur coal.

- Among systems under construction, approximately 75 percent of the equivalent electrical MW capacity is for high sulfur coal application.
- With regard to planned systems, approximately 90 percent of the equivalent electrical MW capacity involves high sulfur coal application.

System Suppliers

There are approximately 30 companies offering FGD systems for utility industry application. Table 2-4 lists the suppliers that have systems either operating or planned for operation along with the number of systems and their equivalent electrical MW capacities. Several firms that have also had systems operating but which are now shutdown are also listed.

Installation Schedules

Schedules for the installation of FGD systems can be highly variable, primarily because of the front end activities relating to process selection and design. In addition, the period from start-up to acceptance by the client can be variable depending upon system performance, and the contractual and other agreements between the utility, its agents, and the system suppliers. The period between contract award and initial system start-up is less variable however. An analysis of the schedules of 35 systems, all calcium based, showed a range of 18 to 60 months with a mean of 31.7 months (2.6 years). The longer lead times are generally associated with new systems that are being constructed as an integral part of a new power generating system.

SECTION 3

OVERALL FGD SYSTEM DEPENDABILITY: TRENDS AND GENERAL CONSIDERATIONS

The dependability of FGD systems as a function of year of start-up date is illustrated in Figure 3-1. This graph shows that the performance of FGD systems, as measured by average operability or availability over total system life, has improved since the early 1970's. Although there is scatter in the results, the correlation is statistically significant. This improved performance is attributable to several general process design and application considerations. Although it is difficult, if not impossible, to quantify the impact of many of these factors, particularly when attempting to apply such an analysis to the entire spectrum of operating FGD systems, we will identify some of the contributing factors and illustrate them, where applicable, with examples. Section 4 will provide more specific information on current technology trends.

Process Design Strategy

Several general tendencies are evident concerning the recent design of FGD systems. Generally the systems have

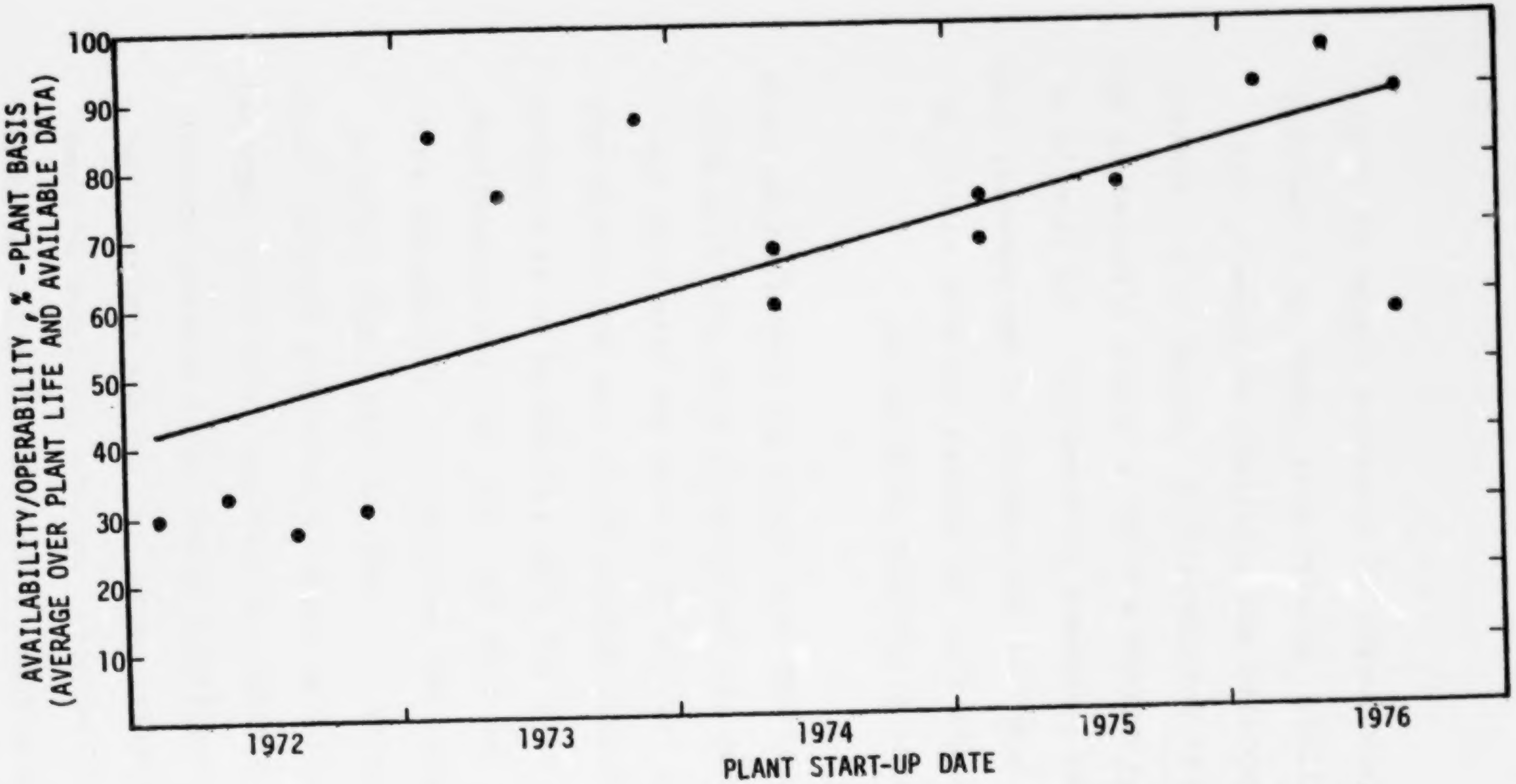

Figure 3-1. Cumulative FGD system dependability (expressed in terms of operability or availability factors) vs. plant startup date.

Notes:
(1) A correlation coefficient for the least squares linear plot of availability/ operability data indicates a statistical certainty of 99%.
(2) Availability or operability was plotted for each system depending upon the available data.
(3) Availability is defined as hours available/hours in time period.
(4) Operability is defined as hours operating/hours boiler operation.

been designed to incorporate an increased degree of flexibility and reliability. Specifically, there is a tendency toward sparing of modules and ancillary components, and toward designing less interdependent systems (i.e., systems where major unit operations are not strongly affected by upstream or downstream component performance). An example of this latter design trend is the removal of the reheater from direct contact with the flue gas stream and thus making is less sensitive to mist eliminator performance.

System Applications

Many of the newer FGD facilities are installed on large baseloaded units which are designed to fire coal from specific sources. This results in a flue gas with more constant and stable characteristics which aids FGD system reliability; the system does not have to respond to as dramatic a variation in flue gas flow rate and flue gas composition. In many of the original FGD applications, the systems were also required to operate on widely varying loads (cycling and peak), and varying coal types (low-sulfur western, high-sulfur eastern, and blends), situations which often required them to respond to conditions beyond their process control capability. Consequently, variations in the reagent feed rate, loss of chemical control, and the incidence of chemically and mechanically related problems resulted in numerous forced outages.

System Supplier and Architectural-Engineer Experience

Later FGD system designs have benefitted from experience gained in the operation of first generation systems. Using this experience, process suppliers and architectural-engineering firms are providing for better process configurations and materials of construction. Indicative of this trend in the fact that many suppliers are now offering systems with broader guarantees covering availability, power consumption, and reagent consumption.

Utility Experience

Along with the increased experience being gained by the system suppliers and architect/engineering firms, utility personnel have been gaining valuable operating and design experience. Several utilities have operated pilot scale FGD systems and have thus been better prepared for operation of full scale systems. Similarly, operation with the first full scale system has led to improved design and operation of subsequent systems. An example of this is the Northern States Power Company (NSP). The design and operation of the scrubbing system installed on Sherburne No. 1 benefitted substantially from previous operating experience gained from a prototype test facility installed at the Black Dog station. Similarly, the full-scale system installed on Sherburne No.

2 approximately one year later, also benefitted substantially from the previous operating experience gained from Sherburne No. 1.

Regulatory Agency Attitudes

As FGD technology has evolved from a research, development, and demonstration (RD&D) effort, enforcement activities by local, state, and federal regulatory agencies have compelled utility companies to improve the reliability of FGD systems.

Process Chemistry

Although chemically related problems (scale, corrosion) are still encountered, and, in some cases, are still very severe, general knowledge concerning their mechanisms of formation and occurance has greatly improved. As a result, systems are being designed and operated so that they will not encounter the same design of problems experienced by the earlier units.

SECTION 4

CURRENT TECHNOLOGICAL TRENDS

Considerable progress has been made in developing both conventional and emerging or advanced FGD processes. Much of this information, which has been acquired from the design and operation of the first generation lime/limestone systems, has been translated into more effective designs and improved operation for the newer systems. These advances are summarized in this section along with a brief overview of the current status of the "emerging processes."

Emerging Processes

Processes within the "emerging" or "advanced" category are defined, somewhat impercisely, as those that incorporate major design and operating changes and thereby differ significantly from conventional direct lime/limestone processes. Of the processes so categorized, a number have been evaluated at pilot and prototype development levels. A few of these systems have progressed to the installation and operation of demonstration units. Table 4-1, provides a brief summary of the emerging processes, and highlights their current level of development and the extent of pre-

Table 4-1. MAJOR EMERGING FGD PROCESSES INVESTIGATED IN THE UNITED STATES

Process	Developer	Current level of development	Previous operating experience	Remarks
Aqueous carbonate	Atomics International	100-MW system (planned)	Mohave pilot plant test program	Full-scale application not yet demonstrated. 100-MW demonstration system scheduled for service in 1978.
Catalytic oxidation	Monsanto	100-MW system (terminated)	Wood River test program	No further process development following unsuccessful demonstration at Wood River.
Chiyoda Thoroughbred 101	Chiyoda International	20-MW system (terminated)	Scholz prototype test program	Development of the process has ceased in favor of a new design concept.
Copper oxide adsorption	Shell/Universal Oil Products	Pilot plant	Big Bend test program	Process available for prototype or demonstration application. No systems planned at present.
Double alkali	A.D. Little, Combustion Equipment	277-MW system (construction)	Scholz prototype test program	Full-scale application not yet demonstrated. A 277-MW demonstration system scheduled for service in 1979.
	FMC	250-MW system (planned)	Industrial systems and utility pilot plants	Full-scale application not yet demonstrated. A 250-MW system is scheduled for service in 1979.
	Buell/Envirotech	575-MW system (planned)	Gadsby pilot plant	Full-scale application not yet demonstrated. A 575-MW system is scheduled in 1979.

Table 4-1 Cont'd).

Process	Developer	Current level of development	Previous operating experience	Remarks
Dry adsorption	Foster Wheeler Bergbau Forschung	20-MW system (terminated)	Scholz prototype test program	No further development of system reported. Further evaluation of the sulfur reduction component (RESOX) scheduled in Germany.
Magnesium oxide	Chemico	150-MW system (terminated) 95-MW system (terminated)	Mystic test program; Dickerson test program	Process demonstrated on full-scale oil- and coal-fired boilers. System now offered for commercial application.
	United Engineers	120-MW system (operational)	Eddystone demonstration test program	Demonstration program now in progress at Eddystone No. 1 of the Philadelphia Electric Co. Pending the outcome of the scheduled one year test program, full-scale application of the process may result at this station and Cromby.
Sodium carbonate	A.D. Little/ Combustion Equipment Associates	Three 115-MW systems (operational)	Reid Gardner Station	Three full-scale sodium carbonate (trona) FGD systems have been in service on coal-fired boilers at the Reid Gardner Station (Nevada Power). System performance has been good. A fourth system is tentatively scheduled for future operation.
Wellman-Lord	Davy Powergas	115-MW system (operational) 375-MW system (construction) 340-MW system (construction)	Crane test program	115-MW NIPSCO/EPA test program now in progress. Two full-scale systems under construction at Public Service Company of New Mexico's San Juan Station. The first San Juan system is expected to begin operation in late 1977.

vious operating experience. As Table 4-1 shows, most of the emerging process experience resides with the sodium or magnesium-based processes. Many of these systems incorporate complex, multiple-loop operations that integrate sulfur dioxide absorption or adsorption, absorbent or adsorbent regeneration, and by-product production into one process. Implicit in Table 4-1 is the lack of long-term operational experience with the nonregenerable processes, which at present, precludes a reliable technical and economic analysis.

Conventional Process

Conventional processes include all direct lime and limestone systems. These systems are the most widely applied and thus are the systems for which there is the most operating experience. Furthermore they are the systems which will have the greatest utilization for at least the near future. For such reasons, these systems have been subjected to extensive analyses. The results of such investigations and general conclusions concerning process design are summarized below.

Reagent Chemistry - To minimize the chemical limitations of lime and limestone systems, efforts have been made to improve the calcium-based process through beneficiation with various compounds. Foremost among these approaches has been the use of magnesium additives. Research conducted at the bench-scale and pilot plant level has stimulated further work on prototype facilities (EPA/TVA Alkali Scrubbing Test Facility, Shawnee Station, Tennessee Valley Authority), and at the demonstration level (EPA Scrubber/Sludge Evaluation Program, Paddys Run Station, Louisville Gas and Electric). Two proprietary absorbents have been developed, one by Dravo (Thiosorbic lime: 2 to 6 percent magnesium oxide lime), the other by Pullman Kellogg (Catalytic limestone: 3 to 27 percent magnesium sulfate limestone). Three full-scale, operational FGD systems (Columbus and Southern Ohio Electric, Conesville No. 5; Pennsylvania Power, Bruce Mansfield No. 1 and No. 2) are now using Dravo's thiosorbic lime reagent. In addition, six more full-scale systems plan to use thiosorbic lime: Big Rivers Rural Electric Power Cooperative, Reid No. 2; Columbus and Southern Ohio Electric, Conesville No. 6; Duquesne Light, Elrama Nos. 1-4; Duquesne Light, Phillips Nos. 1-6; Indianapolis Power and Light, Petersburg No. 3; and Pennsylvania Power, Bruce Mansfield No. 3. To date, there are no plans to use Pullman Kellogg's catalytic limestone.

Another important development in reagent chemistry is the use of fly ash alkalinity as the principal reagent.

Many scrubbers installed in the western United States for particulate control only, were removing appreciable amounts of sulfur dioxide (30 to 40 percent) and developing sulfate scale without the use of external reagents. (Sulfur dioxide removal occurs because of the inherent alkalinity of the fly ash associated with western lignite and subbituminous coals. Many western coal ashes contain high concentrations of sodium oxide, magnesium oxide, and, in particular, calcium oxide. The alkali content tends to be lower for higher ranked coals). Research into the feasibility of alkali fly ash scrubbing for sulfur dioxide removal has been conducted since 1971 by the Grand Forks Energy Research Center of the U.S. Energy Research and Development Administration (ERDA). Four separate pilot plant programs have resulted in the development of lime and limestone scrubbing processes that use the alkaline content of the fly ash as the primary reagent source, and use supplemental lime and limestone for pH and scale control. Currently, seven full-scale alkaline fly ash/lime and alkaline fly ash/limestone systems are in service: Kansas City Power and Light, Hawthorn No. 3 and No. 4; Minnkota Power Co-op., M.R. Young No. 2; Montana Power, Colstrip No. 1 and No. 2; and Northern States Power, Sherburne No. 1 and No. 2. Five more systems are planned for future operations: Montana Power, Colstrip No. 3 and No.

4; Northern States Power, Sherburne No. 3 and No. 4; and United Power Association, Coal Creek No. 1 and No. 2.

Particulate Precollection - Particulate precollection using an upstream electrostatic precipitator or baghouse minimizes a number of chemical and mechanical problems associated with simultaneous or two-stage wet scrubbing systems. These include minimizing corrosion at wet/dry interface areas and gypsum scale formation within the FGD system; some metallic components in the fly ash provide catalytic oxidation sites for the promotion of sulfite to sulfate oxidation in the scrubbing solution, thus resulting in sulfate saturation and gypsum scale development. In addition, erosion is reduced substantially. Beneficial design aspects include the capacity to use dry, forced-draft booster fans with less expensive materials of construction.

FGD Booster Fans - If particulate precollection is reliable, the booster fan may be placed upstream of the FGD system, since the erosive nature of hot, fly ash-laden flue gas will have been largely eliminated. Such a design improves overall system dependability. Many of the booster fans on the initial FGD systems operated on saturated gas resulting in acid attack, corrosion, solids deposition erosion, and scale development on the fan blades and housing, which often led to forced outages.

Dampers - Efficient and reliable gas dampers allow modules to be maintained during by-pass situations, without the necessity of unit shutdown. This can minimize the outage time assessed against the FGD system and thus improve overall system dependability. Corrosion and erosion of various types of dampers have been widely experienced. Promising results with seal air components (e.g. double lower seal air dampers) have been reported in recent operations and are incorporated into the design of some future systems.

Presaturator - Cooling the flue gas to its adiabatic saturation temperature, prior to contact with the scrubbing slurry, increases sulfur dioxide removal efficiency and minimizes the potential for corrosion and scaling at the slurry/gas interface areas. Presaturators were not incorporated in the design of many of the initial FGD systems which experienced the chemical problems mentioned above. This situation was due, in part, to the incorporation of venturi scrubbers for wet phase particulate removal in the FGD system. Presaturators are now used, or are being planned for use, in those systems that include dry phase particulate precollection.

Absorbers - Spray towers have a minimum of internal components in contact with the flue gas stream. They offer the potential for higher availabilities, because of the lack of sites for deposition of solids in the form of scale products, unused reagent, and uncollected fly ash. In many of the initial FGD installations, the absorber modules used packed towers or other designs which had internal sites to promote contact between the flue gas and alkaline slurry. The major drawback with such towers is that the internals are prone to plugging and scaling, and require higher maintenance. In many cases this maintenance necessitates shutdowns for clean-out, thereby reducing overall system availability. In lieu of potentially trouble-some internal sites, spray towers are used with high liquid-to-gas ratios (L/G), between 60 and 100 gpm per 1000 cfm.

All major system suppliers now offer a spray tower design. Both vertical and horizontal spray towers are available.

Mist Elimination - Chevron and baffle-type mist eliminators have been and are currently being used in virtually every FGD system installed in the United States. Although a number of different designs have been tested (wire mesh, tube bank, gull wing, electrostatic precipitator, radial vane), emphasis on the use of baffle and chevron designs will continue for future systems. The popularity of these collectors is due primarily to design simplicity, excellent collection efficiency (for moderate to large size drops), low pressure drop, wide-open construction, and low cost. Within these two preferred types of mist eliminators, a number of specific design and construction trends have been noted:

- Chevron designs (continuous vane construction) predominate over baffle designs (discontinuous slat construction).
- Fiberglass-reinforced plastic has become the predominant construction material over stainless steel (316L) and other types of plastics.
- The horizontal configuration (vertical gas flow) is preferred to vertical configurations (horizontal gas flow).
- Two-stage designs predominate over single stage designs.
- Bulk entrainment separators, perforated plates, impingement plates, and other precollection devices are becoming integral parts of mist elimination systems.
- Wash water trays are being incorporated into the mist elimination systems more frequently, in order to maximize fresh wash water usage.
- Three-pass designs predominate over two-pass designs. Where prior stages or precollectors are incorporated, two-pass designs predominate.
- Mist eliminator wash systems that employ intermittent, high-velocity sprays predominate over continous wash systems.

Reheat - Virtually all the FGD systems coming on line and planned for future operation incorporate some type of stack gas reheat system to avoid condensation and corrosion to downstream equipment, ductwork, and stack. Suppresses plume visibility and enhances plume rise and pollutant dispersion. To date, a number of "wet stack" FGD systems (no reheat) have been installed and have encountered corrosion problems. The trend in reheat strategies is toward indirect hot air reheat and flue gas by-pass reheat, and away from in-line reheat and direct combustion reheat. In-line reheat systems have been subject to corrosion and solids deposition, the latter occurring because of inefficient upstream mist elimination. This has often resulted in reduced heat transfer, leading to subsequent corrosion of downstream equipment. Direct combustion methods require natural gas or fuel oil. Of the two preferred strategies, by-pass reheat is the predominate method for low-sulfur, western coal FGD applications, where the maximum degree of reheat is not seriously constrained by pollution emission standards. Indirect hot air reheat methods predominate for those FGD systems that treat medium- to high-sulfur coal flue gas.

Recirculation Pumps - Recirculation pumps are the largest and most expensive pumps in the lime and limestone slurry systems. Because of their critical importance to system operation, a number of special considerations are involved in their design. Specifically, special attention must be given to an accurate service description (solution pH, specific gravity, fly ash content, gas entrainment), and to flow rates, head, impellers, drives, seals, construction materials, and redundancy. Because of the number of site-specific variables that must be considered, an accurate analysis of trends in the design and operation of recirculation pumps is not possible. A number of general trends are evident, however, and are summarized below:

- All systems now incorporate spare pumps. Spare capacity generally ranges from 50 percent (one spare for every two operating pumps) to 100 percent (one spare per operating pump).
- Natural and synthetic molded rubber lining is specified for wetted parts in the majority of the pumps.
- Flush water wash systems are provided to purge the pumps of solids, which tend to settle out during periods of inactivity.
- Higher flow rate pumps (15,000 to 20,000 gpm) predominate over lower flow rate pumps (6,000 to 10,000 gpm), because of the trend toward high L/G spray towers.

Solids Separation - A number of solids dewatering processes have been used both experimentally and commercially for flue gas cleaning wastes generated by lime and limestone systems. Four techniques have been demonstrated at the pilot, prototype, and full-scale developmental levels: interim ponding, clarification, centrifugation, and vacuum filtration. A number of specific trends are evident. The major items in this regard, however, are the increasing of emphasis being placed on clarification, centrifugation, and vacuum filtration techniques, and the decreasing emphasis on interim ponding. Formerly, an interim pond was relied on to fulfill three functions: clarification, dewatering, and temporary or final sludge storage. The realization that a single pond cannot perform all three functions has spurred the development of the other techniques. Furthermore, increasing emphasis on off-site disposal for landfill and structural fills, plus increased emphasis on attaining closed water loop operations, has also stimulated the use of clarifiers, centrifuges, and vacuum filters. In addition to these techniques, a number of installations are experimenting with or using forced oxidation strategies to enhance solids settling properties, to decrease sludge disposal land requirements, and to maximize the quality of recycled water. Two full-scale systems using forced oxidation are now in service (Northern States, Sherburne No. 1 and No. 2), and two additional full-scale systems are using forced oxidation on a short-term experimental basis (Arizona Public Service, Cholla No. 1; and Commonwealth Edison, Will County No. 1). Two additional installations are planning to use forced oxidation for future operations (Northern States Power, Sherburne No. 3 and No. 3), and two pilot plants are actively involved in forced oxidation test programs (TVA/EPA Alkali Scrubbing Test Facility, Shawnee No. 10, Tennessee Valley Authority, and the EPA/IERL pilot plant, Research Triangle Park).

Sludge Disposal - A major problem associated with the non-regenerable lime and limestone systems is the tremendous amount of waste material that must be handled and disposed of in an environmentally acceptable manner. For disposal methods to be environmentally acceptable, regardless of the specific processes employed, they must be capable of preventing or minimizing the seepage of waste liquors carrying potentially toxic compounds into water supplies. On-site ponding and landfilling with physically conditioned or chemically treated waste come into this acceptable category. A number of specific disposal methods are available. The selection of a method depends on a number of site-specific factors. However, some general developments and trends have been identified.

- By the end of 1976, approximately 60 percent of the FGD operating capacity was using on-site ponding for the ultimate disposal of flue gas cleaning wastes. The remaining FGD operating capacity was using chemical fixation and physical conditioning methods for landfill disposal.
- Based upon the strategies now in service or planned for future operations, approximately 75 percent of the FGD operating capacity will be

using on-site ponding for ultimate disposal of flue gas cleaning wastes. The remaining 25 percent capacity will employ physical and chemical treatment methods for landfill disposal.

- Concerning chemical stabilization methods, two major suppliers now offer proprietary methods for chemically fixation of flue gas cleaning wastes. The Dravo Corporation and IU Conversion Systems (IUCS) both market commercial stabilization processes.
- A number of utilities operate FGD systems that include fly ash and lime stabilization of the thickener underflow or vacuum filtered solids prior to off-site landfill.
- A number of major system suppliers now offer sludge treatment processes in addition to emission control systems. Research-Cottrell and Combustion Engineering, for example, now provide sludge treatment systems as separate packages from the FGD systems.

Construction Materials - Analysis of FGD experience in the United States to date, when examined at all levels of development, makes one point clear: There are so many factors and so many apparent contradictions in FGD operation that generalization is difficult. Nowhere in FGD technology is this more evident than in the area of construction materials. Many examples can be cited where seemingly inferior construction materials have been adequate, whereas apparently adequate materials have failed. Sufficient data have been accumulated, however, to provide a general trend analysis for the construction of critical elements in FGD systems.

- 316L stainless steel predominates as the preferred material of construction for the FGD modules, inlet and outlet ducts, and all other wet surfaces. The preference for 316L stainless steel is based on its superior resistance to corrosion, erosion, and scaling. Information and data from previous and current operational FGD systems, system suppliers, architectural-engineering firms, and independent test and evaluation programs, have shown that carbon steel, low alloy steels, type 304, and type 304L are inadequate in regard to corrosion resistance. Superior erosion resistance of 316L, when evaluated with the above-mentioned metallic materials, and even with non-metallic coatings, has been demonstrated for slurry and flue gas ash-laden service. Also, some information indicates that scale has less tendency to adhere to surfaces constructed of 316L stainless steel, and is easier to clean. The successful application of 316L stainless steel has been attributed to the molybdenum content of the metal, resulting in a number of major system suppliers specifying a minimum of 2.50 to 2.75 percent molybdemun for all wetted surfaces.

- Synthetic and natural rubber coatings predominate in recycle tanks, pumps, and lines. These materials have been reported to give superior erosion resistance once application problems have been overcome.

- Some systems are also incorporating such alloys as Hastelloy C-276, Hastelloy G, Inconel 625, Incoloy 825, 317L stainless steel, 904L stainless steel and Jessop JS700 in wet/dry high temperature, high chloride environments, such as in presaturators.

- Regarding the use of liners in the absorbers, exhaust ducts and stacks, a number of materials such as resins, ceramics, polyesters, polyvinyls, polyurethanes, carboline, and Gunite, have been used with varying degrees of success. Although successful applications have been reported, widespread failures of the liners have been attributed to the unvelreliability and inexperience of lining applicators, instability of the materials at high temperatures, and inconvenience of repair.

STATUS OF SO_2 AND NO_x REMOVAL SYSTEMS IN JAPAN

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ABSTRACT

Major FGD installations being operated in Japan consist of 140 gypsum by-producing plants with a total capacity of 18,000 MW equivalent and 25 plants (4,000 MW equivalent) by-producing sulfuric acid, sulfur, and ammonium sulfate. In addition, there are about 800 smaller plants (9,000 MW equivalent) by-producing sodium sulfite and sulfate. Not many FGD plants will be installed in the future, partly because ambient SO_2 concentration has almost attained the low level required by stringent regulations and partly because the by-products are now in a state of oversupply.

On the other hand, many commercial NO_x removal plants have been put into operation or are under construction chiefly using selective catalytic reduction. The combination of the wet FGD and dry NO_x removal systems, however, has presented problems. Several processes for simultaneous removal of SO_2 and NO_x have been developed to avoid these problems.

The present paper will describe some operation data of 23 major FGD plants and will outline a few new FGD and simultaneous removal processes.

STATUS OF SO_2 AND NO_X REMOVAL SYSTEMS IN JAPAN

1. MAJOR FGD PROCESSES AND PLANTS

Table 1 lists major constructors of FGD plants and numbers and capacities of plants operational at end 1977. The plants totaled more than 500 and their combined capacity reached 85,000,000 Nm^3/hr (equivalent to 28,000 MW). About a half of this capacity is accounted for by utility boilers (mostly oil-fired) and the rest by industrial boilers, iron-ore sintering machines, nonferrous metal industry, sulfuric acid plants, etc.

About 50% of all the plants, in terms of capacity, use the wet lime/limestone process to by-produce gypsum, 16% the indirect lime/limestone process (double alkali type), 13% regenerable processes to by-produce sulfuric acid, ammonium sulfate and elemental sulfur, and 24% sodium scrubbing to by-produce sodium sulfite or sulfate. The average plant capacity is 427,000 Nm^3/hr for the wet lime/limestone, 291,000 Nm^3/hr for the indirect lime/limestone, 378,000 Nm^3/hr for the regenerable processes, and 59,600 Nm^3/hr for the sodium scrubbing processes. About 80% of the sodium scrubbing plants by-produce sodium sulfite for paper mills and the rest oxidize the sulfite by air bubbling to sulfate, which is either used in the glass industry or purged in wastewater.

In addition to the 335 sodium scrubbing plants listed in Table 1, there are nearly 500 smaller ones operated commercially with an average capacity of about 20,000 Nm^3/hr.

Table 1. NUMBERS AND CAPACITIES (1,000 Nm^3/hr) OF FGD PLANTS BY MAJOR CONSTRUCTORS (OPERATIONAL AT END 1977)

Plant constructor	Wet lime limestone	Indirect lime limestone	H_2SO_4, S $(NH_4)_2SO_4$	Na_2SO_3 Na_2SO_4	Total
Mitsubishi Heavy Industries (MHI)	33 (18,270)			3 (292)	36 (18,562)
Ishikawajima H.I. (IHI)	17 (4,445)			79 (4,351)	96 (8,796)
Hitachi, Ltd.	13 (6,940)*		2 (590)	15 (603)	30 (8,133)
Mitsubishi Kakoki (MKK)	2 (256)		13 (6,478)**	41 (913)	56 (7,643)
Kawasaki Heavy Industries	4 (756)	6 (5,450)		7 (256)	17 (6,380)
Tsukishima Kikai (TSK)	1	4 (398)	1 (88)	40 (4,042)	45 (4,608)
Chiyoda Chemical Eng. & Construc.		14 (4,459)			14 (4,459)
Oji Koei				57 (4,280)	57 (4,280)
Fuji Kasui Engineering	7 (3,954)			6 (270)	13 (4,224)
Kurabo Engineering		5 (413)	1 (18)	106 (3,751)	112 (4,182)
Mitsui Miike-Chemico	4 (2,744)		1 (500)		5 (3,244)
Ebara Manufacturing		11 (1,914)		10 (1,167)	21 (3,081)
Nippon Kokan (NKK)	3 (245)	1 (150)	2 (1,990)	6 (62)	12 (2,447)
Kureha Chemical				8 (1,431)	8 (1,431)
Showa Denko				5 (1,372)	5 (1,372)
Gadelius				8 (1,291)	8 (1,291)
Sumitomo (SCEC)-Wellman			6 (1,288)		6 (1,288)
Mitsui Metal Engineering	4 (1,006)		2 (130)		6 (1,136)
Kobe Steel	5 (1,125)				5 (1,125)
Japan Gasoline	1 (330)		1 (125)		2 (455)
Dowa Engineering		5 (453)			5 (453)
Niigata Iron Works		1 (185)			1 (185)
Mitsui Shipbuilding				1 (160)	1 (160)
Sumitomo Heavy Industries			1 (150)		1 (150)
Total	94 (40,181)	47(13,422)	30(11,357)	335(19,961)	506 (84,921)

* Babcock - Hitachi　　** Wellman - MKK

Tables 2 - 4 show operation data of major FGD plants. The data of the plants owned by the electric power companies -- Chubu, Chugoku, Shikoku, Kyushu, Hokuriku, Tokyo and Electric Power Development Co. -- Mitsui Aluminum, and Idemitsu Kosan were supplied by the plant owners while other data were made available by the process developers.

Wet lime/limestone process plants shown in Table 2 by-produce salable gypsum except the Omuta plant, Mitsui Aluminum, which by-produces a throw-away sludge. For the production of gypsum, a calcium sulfite sludge at a pH of about 4 is air-oxidized. In order to lower to 4 the pH of the slurry discharged from the scrubber, an additional scrubber is used at the Owase plant, Chugoku Electric (MHI process), and the Takasago plant, Electric Power Development Co. (Mitsui-Chemico process), while sulfuric acid is used at other plants.

SO_2 removal efficiency ranges from 90-98%, and power required for a total FGD system ranges from 2.0-3.5% of the power generated; power requirement is larger for a higher SO_2 removal efficiency because of the larger pressure drop of the gas in the scrubber to attain the higher removal efficiency.

Wastewater is purged at the rate of 3-30 tons/hr or 8-60 kg/MWhr, primarily to prevent the accumulation of chlorine in the scrubber liquor because chlorine increases corrosion. The plants of Electric Power Development Co. give much wastewater (60 kg/MWhr) due to the use of coal which contains a considerable amount of chlorine, while two plants for an oil-fired flue gas purges only 8 kg/MWhr. The ratio of water to by-product gypsum is about 0.2 for oil-fired flue gas and 1.5 for coal-fired flue gas. Since the by-product gypsum contains about 10% moisture, the amount of water removed from the system in the ratio of 1.0 is almost equal to that in a throw-away process of a sludge with 40% solid without wastewater.

The Omuta plant, Mitsui Aluminum, has been operated with a low oxidation ratio without the formation of gypsum while all the other plants have been operated with forced oxidation with a considerable amount of gypsum crystals in a circulating slurry; in both cases, scaling is prevented. The Omuta plant and the Tamashima plant, Chugoku Electric, have a stand-by scrubber which has been proved virtually unnecessary through operating experience. Other plants have no stand-by scrubber.

Some of the plants encountered problems at start-up but most of the problems were solved in a few months and since then all of the plants have attained an operability better than 95%. Operability means an FGD plant's operating hours per cent of the scheduled operating hours of the gas source -- normally 11 months' continuous operation with a month's shutdown for utility boilers and about three months' continuous operation followed by several days' shutdown for sintering machines. Although most FGD plants are not entirely free from such problems as scaling and corrosion, those can be controlled so that they do not hinder operation and are removed during scheduled shutdown of the gas source.

Table 2. OPERATION DATA OF MAJOR LIME/LIMESTONE PROCESS PLANTS

Process	Lime scrubbing		Limestone scrubbing					
Process developer	Mitsubishi H. I.	Chemico-Mitsui	Mitsubishi H. I.	Babcock-Hitachi	Babcock-Hitachi	Mitsui-Chemico	IHI-Chemico	Sumitomo-Fujikasui
Plant owner	Chubu Electric	Mitsui Aluminum	Kyushu Electric	Chugoku Electric	Electric Power D.C.	Electric Power D.C.	Electric Power D.C.	Sumitomo Metal
Plant site	Owase	Omuta[a)]	Karatsu	Tamashima	Takehara	Takasago	Isogo	Kashima
Fuel	Oil	Coal	Oil	Oil	Coal	Coal	Coal	Coke
FGD capacity (MW)	375	156	250	500	250	250	265	(630)[b)]
FGD capacity (1,000 Nm^3/hr)	1,200	510	730	1,480	850	850	900	2,000
Inlet SO$_2$ (ppm)	1,600	2,300	530	1,460	1,700	1,500	400	400-600
Inlet dust (mg/Nm^3)	10	630	25	40	600	100	1,500	100-200
Inlet gas temperature (°C)	150	135	150	140	150	140	170	150
CaO/SO$_2$ stoichiometry	1.0	1.05			1.05	1.0	1.0-1.05	1.05
Number of scrubbers in parallel	2	1+1[c)]	1	3+1[c)]	1	1	1	2
Prescrubber (first scrubber)								
Type	Spray	Venturi	Spray	Venturi	Venturi	Venturi	Venturi	pp[d)]
L/G (liters/Nm^3)	2	5-8	2	10	2.4	6	7	7
Scrubber (second scrubber)								
Type	Packed	Venturi	Packed	pp[d)]	pp[d)]	Venturi	Venturi	pp[d)]
Slurry pH	6.5-7	7.5	6.2	6.6	6-6.5	6.2	5-6	6.7
Slurry concentration (%)	10	5		12.5	9	5-6	7	7-8
L/G (liters/Nm^3)	7	5-8	12	10	7	6	7	8
Gas velocity (m/sec)	3.4		3.0				3.0	4.4
Outlet SO$_2$ (ppm)	120	220	50	60	100	100	30	30
Outlet dust (mg/Nm^3)	10	40	6		30	50	50	20
SO$_2$ removal efficiency (%)	93	90	90	96	98	93	93	93-95
Mist eliminator type	C·E[e)]	Chevron	Chevron	Pwf[f)]	Pwf[f)]	Chevron	Chevron	Impinger
Pressure drop (mmH$_2$O) Prescrubber	30	} 200	60	225	230	150	150	180
Scrubber	150		35	505	375	150	150	130
Mist eliminator	25	30	65	25	10	25	50	30
Total system	375		185	1,080	950	525	600	530
Wastewater purged (t/hr)	3			4	15	10	15	30
Energy requirement (design)								
Pump (kW)		1,320		4,100	1,000	} 5,670	2,000	4,400
Fan (kW)		2,000		11,500	5,500		3,600	3,300
Total FGD system (kW)	7,500	3,867	5,070	17,600	7,800	7,500		
Per cent of power generated	2.0	2.5	2.0	3.5	3.1	3.0	2.1[g)]	1.2[g)]
Operability (%)	98	100	100	98	Above 95	97	Above 99	100

a) Throw-away process. Others by-produce gypsum. b) Iron-ore sintering plant. Others are utility boilers.
c) Stand-by. d) Perforated plate. e) Chevron and Euraform. f) Pipe with fin. g) For pump and fan.

As is generally experienced, mist eliminators are most susceptible to scaling. Scaling in this case is thought to occur for the most part as lime or limestone in the mist reacts with SO_2 and O_2 in the gas, to form gypsum on the wall of the eliminator. It seems that the scaling problem is less in Japan than in the U.S.A. This may be attributable to the following reasons:

(1) A low concentration of lime or limestone in the mist due to a high utilization (over 90%) of lime or limestone; CaO/SO_2 mole ratio of 1.0-1.05 is used to remove 90-98% of SO_2.

(2) A low concentration of SO_2 in the gas passing through the eliminator which is due mainly to the high SO_2 removal ratio.

(3) Mist and wash liquor contain a considerable amount of gypsum which works as crystal seed to prevent scaling.

(4) The eliminator is usually washed with a circulated liquor but when an appreciable amount of scale is formed, it is washed with fresh water to dissolve the scale.

A lime/limestone process plant costs $45-65/kW in battery limits at current prices. Since a major portion of desulfurization cost derives from fixed costs, efforts have been made to lower capital cost.

Table 3 shows the operation data of major plants by-producing gypsum by other processes. Most of the processes are indirect lime/limestone processes except the Kobe Steel and Kawasaki processes which use -- in addition to lime -- calcium chloride and magnesia, respectively. Those two processes will be described in 2.1 and 2.2.

The pH of the absorbing liquor of the indirect processes ranges from 1.0 (Chiyoda) to 6.8 (Showa Denko) and the L/G is smaller with processes with a higher pH. Generally speaking, energy requirement is smaller with processes with a higher pH which gives a smaller L/G. On the other hand, operability of the plant is higher with a lower pH. A low pH and the use of a large L/G are favored for scale prevention.

For processes which use a liquor with pH below 4.0 (Chiyoda, Dowa, and Kurabo), air oxidation is carried out with the liquor and proceeds more readily than with the calcium sulfite in the other processes.

Generally speaking, the wet lime/limestone processes (direct processes) in Table 2 and the processes in Table 3 are just about equal in SO_2 removal efficiency, power consumption, and operability. Investment cost of the plants ranges from $45-80/kW in battery limits at current prices. All of the plants using an absorbing liquor with a pH below 5.5 have over 99% operability. Such a process may suit plants in which precise operation control is difficult.

Table 4 shows some data of regenerable process plants. Nippon Kokan recently started up two large plants by-producing ammonium sulfate, which are described in 2.3.

Table 3. OPERATION DATA OF OTHER GYPSUM BY-PRODUCING WET PROCESS PLANTS

Process developer	Kureha-Kawasaki	Showa Denko	Chiyoda	Dowa	Kurabo	Nippon Kokan	Kobe Steel	Kawasaki H.I.
Absorbent	NaOH	NaOH	H_2SO_4	$Al_2(SO_4)_3$	NH_3	NH_3	$CaO-CaCl_2$	$Mg(OH)_2$
Precipitant	$CaCO_3$	$CaCO_3$	$CaCO_3$	$CaCO_3$	CaO	CaO		CaO
Plant owner	Shikoku Electric	Showa Denko	Hokuriku Electric	Naikai Salt		Nippon Kokan	Nakayama Steel	Unitika
Plant site	Sakaide	Ichihara	Fukui	Tamano		Keihin	Funamachi	Okazaki
Fuel	Oil	Oil	Oil	Oil	Oil	Coke[a]	Coke[a]	Oil
FGD capacity (1,000 Nm^3/hr)	1,270	500	980	82	115	150	375	200
FGD capacity (MW)	450	170	350	(30)	(40)	(50)	(130)	68
Inlet SO_2 (ppm)	1,270	1,400	1,800	1,500	1,480	350	150-250	1,400
Inlet dust (mg/Nm^3)	20	100-200	30	20	150	80	300-400	200
Inlet gas temperature (°C)	135	140	140	170	170	120	140-155	170
Number of scrubbers in parallel	2	4	1	1	1	1	1	1
Prescrubber type	None	None	Venturi	Spray		Spray	Spray	None
L/G ($liters/Nm^3$)				2.5	1.4	1.0	3.5	
Scrubber type	Packed	VC[b]	Packed	Packed	Packed	Screen	Venturi	MV[c]
Liquor pH	6.2	6.8	1	3.5	3.8	6.0	5-5.5	5-6
Concentration	20	25	1-2		10	30	30+6[d]	
L/G ($liters/Nm^3$)	1.2	0.5-1	40-60	10	8	2	3	6
Gas velocity (m/sec.)	2			0.9	2	2	3	3
Outlet SO_2 (ppm)	70	75-100	80	100	130	10-20	15-25	Below 140
Outlet dust (mg/Nm^3)	10	Below 50		10	Below 50		40-50	Below 100
SO_2 removal efficiency (%)	95	90-95	96	93	91		90	Above 90
Mist eliminator type	Terellette		Euraform	Wire mesh	Euraform		Euraform	Louver
Pressure drop (mmH_2O) Prescrubber				10-20	130		15	
Pressure drop (mmH_2O) Scrubber	115	300-500		70-80	120		120	115
Pressure drop (mmH_2O) Mist eliminator	25	30-50		20-40			35	50
Pressure drop (mmH_2O) Total system	310	400-700	680	170	300	300	220	230
Wastewater purged (t/hr)	None	4-6	7-24	0.03	None	2	None	None
Energy requirements (Design)								
Pump (kW)			5,300	150	470		1,425	430
Fan (kW)			5,500	300	190		1,000	370
Total system (kW)	7,900		12,300	590	880	400	3,200	1,160
Per cent of power generated	1.8		3.5	2.0	2.2	0.8	2.5	1.7
Operability (%)	95	98	100	Above 99	Above 99	Above 95	Above 99	99

a) Iron-ore sintering plant b) Vertical cone c) Multi-venturi d) 30% $CaCl_2$ + 6% CaO

Wellman-Lord process plants have been operated smoothly but need extensive wastewater treatment including ozone oxidation to decompose polythionates such as $Na_2S_2O_6$, which form mainly at the heating step of the absorbing liquor. It seems that polythionates are formed also in other wet processes, even though in small amounts, and might necessitate treatment when wastewater regulations are tightened.

The Chemico-Mitsui magnesium scrubbing plant of Idemitsu was beset by problems mainly in regeneration steps for over one year after its start-up in 1975. The problems have since been solved through the efforts of Idemitsu and Mitsui Miike. Normally no wastewater is purged from the system.

The Shell process plant of SYS encountered problems, most of which have been solved. Further improvements are on the way, to achieve a higher operability. The process seems costly, requiring as it does large reactors and hydrogen. The capability of simultaneous removal of NO_x may compensate for the disadvantage (3.2.4).

The Kashima plant, Tokyo Electric, using carbon absorption and water wash, has been operated for nearly 5 years without appreciable problems. Carbon consumption has proved very low (about 2% yearly) owing to the use of a fixed bed and to regeneration by water-wash. (The Sakai plant, Kansai Electric, which used a carbon process with a moving bed and thermal regeneration, was shut down recently because of high consumption of carbon.) The Kashima plant by-produces a dilute sulfuric acid (about 17%), which is treated with powdered limestone to produce salable gypsum of good quality. Hitachi Ltd. recently constructed another carbon process unit with a capacity of treating 150,000 Nm^3/hr of flue gas from an oil-fired boiler for Unitika Co. at the Uji plant. The dilute sulfuric acid obtained by water-wash is sprayed into the incoming flue gas and, concentrated to 60%, is used in the Uji plant. Sulfuric acid mist formed in the concentration step is caught by the carbon bed and causes no problem.

The investment costs for the sulfuric acid and sulfur by-producing plants excluding the Claus furnace range from $80-130/kW in battery limits at current prices.

Although most of the regenerable process plants as well as other process plants have achieved a high operability, not many new FGD plants will be constructed due to the oversupply of by-products.

Table 4. OPERATION DATA OF REGENERABLE PROCESS PLANTS

Process developer	Nippon Kokan	Wellman-MKK	Wellman-SCEC	Chemico-Mitsui	Kureha	Shell^c)	TEPCO-Hitachi^c)
Absorbent	NH_3	NaOH	NaOH	MgO	NaOH	CuO	Carbon
By-product	$(NH_4)_2SO_4$	$SO_2 \rightarrow H_2SO_4$	$SO_2 \longrightarrow H_2SO_4$	$SO_2 \rightarrow S$	Na_2SO_3	$SO_2 \rightarrow S$	$H_2SO_4 \rightarrow CaSO_4$
Plant owner	Nippon Kokan	Chubu Electric	Sumitomo Chemical	Idemitsu Kosan	Mitsui Toatsu	Showa Y.S.	Tokyo Electric
Plant site	Keihin	Nishinagoya	Sodegaura	Chiba	Nagoya	Yokkaichi	Kashima
Fuel	Coke^a)	Oil	Oil	Oil	Oil	Oil	Oil
FGD capacity (1,000 Nm^3/hr)	1,120	620	370	460^b)	190	116	420
FGD capacity (MW)	(380)	220	130	(160)	65	38	150
Inlet SO_2 (ppm)	350	1,600	1,500	2,850	1,400	1,250	150
Inlet dust (mg/Nm^3)	50		100		200-300	Below 50	
Inlet gas temperature (°C)	120	140	160	185	170		130
Number of scrubbers in parallel	2	1		1		1^d)	
Prescrubber type	Spray			Venturi			
L/G (liters/Nm^3)	1.0						
Scrubber type	Screen	Sieve tray		Venturi	Packed	PP^e)	Packed
Liquor pH	6.0				6.5		
L/G (liters/Nm^3)	1.0	0.6			1.2		
Gas velocity (m/sec)	1.6	1.8			Below 2.0		
Outlet SO_2 (ppm)	10-20	120	Below 150	120	6	125	30
Outlet dust (mg/Nm^3)	10	35	Below 50			Below 50	
SO_2 removal efficiency (%)	94-97	92	Over 90	95		90	80
Mist eliminator type	Wet EP			Chemico			
Pressure drop (mmH_2O): Prescrubber							
Pressure drop (mmH_2O): Scrubber		400			165	200	630
Pressure drop (mmH_2O): Mist eliminator		50			40		
Pressure drop (mmH_2O): Total system	250	550		500^f)	250	400-500^f)	870
Wastewater purged (t/hr)	10	4	Some	0.1	Some	Some	13
Energy requirements (Design)							
Pump (kW)		840		1,960		140	280
Fan (kW)		2,350		3,400		730	2,700
Total FGD system (kW)	1,950				560	870^f)	3,245
Per cent of power generated	0.5	1.5^g)		3.4^g)	0.9	2.3^f)	2.2
Operability (%)	100	99.5	Over 95	100	100		92

a) Iron ore sintering plant b) From oil burner and Claus furnace c) Dry process d) Two reactors are used alternately for SO_2 absorption and regeneration e) Parallel passage reactor f) Excluding Claus furnace g) For pump and fan

2. NEW FGD PROCESSES AND PLANTS

2.1 Kobe Steel Process

Three plants by the Kobe Steel process are in operation using a 30% calcium chloride solution containing CaO as the absorbent. Operation data of one of the plants are shown in Table 3. Flue gas is cooled by a dilute $CaCl_2$ solution discharged from a gypsum centrifuge and is then treated in a special type of scrubber (combination of spray and venturi) by an absorbent containing 30% $CaCl_2$ with dissolved lime and 6% solids (Figure 1). Lime is much more soluble in the $CaCl_2$ solution than in water. The slurry leaving the scrubber is centrifuged and the calcium sulfite sludge is repulped with water and a small amount of sulfuric acid, then oxidized by air to form gypsum. The liquor from the centrifuge is returned to the cooler, giving no wastewater at all.

Since the solubility as well as degree of supersaturation of calcium sulfate in the $CaCl_2$ solution is about 1/10 of that in water, less scaling is expected. The vapor pressure of the liquor is low and the gas temperature after the scrubbing reaches 70°C as compared with the 55-60°C for other wet FGD processes. Thus, less energy is required for reheating.

In the early period of operation, when the solid concentration of the slurry in the scrubber was about 1%, slight scaling occurred in the piping of the scrubber system. The solid concentration was later increased to 5-6% and since then no scaling problem has occurred for about a year.

The process is of interest because in any gypsum by-producing plant for flue gas from a coal-fired boiler, chlorine would accumulate in the scrubber liquor to give an about 30% $CaCl_2$ solution unless water in addition to the approximately 10% moisture in by-product gypsum is purged from the system.

2.2 Kawasaki Magnesium Gypsum Process

It has been known that addition of a small amount of MgO to a lime/limestone scrubbing liquor helps prevent scaling. The Kawasaki process uses a fairly high concentration of MgO (about 5% in the liquor) which works also as an absorbent (Figure 2). Operation data of one of the two plants are shown in Table 3. A multi-venturi type scrubber is used without a cooler. The slurry discharged from the scrubber is oxidized by air, then is centrifuged to by-produce salable gypsum. The liquor discharged from the centrifuge is returned to the scrubber, giving no wastewater. Major reactions are shown below:

$$Mg(OH)_2 + SO_2 = MgSO_3 + H_2O$$
$$MgSO_3 + SO_2 + H_2O = Mg(HSO_3)_2$$
$$Mg(HSO_3)_2 + CaO = MgSO_3 + CaSO_3 + H_2O$$
$$CaSO_3 + 1/2O_2 = CaSO_4$$
$$MgSO_3 + 1/2O_2 = MgSO_4$$
$$MgSO_4 + CaO + H_2O = Mg(OH)_2 + CaSO_4$$

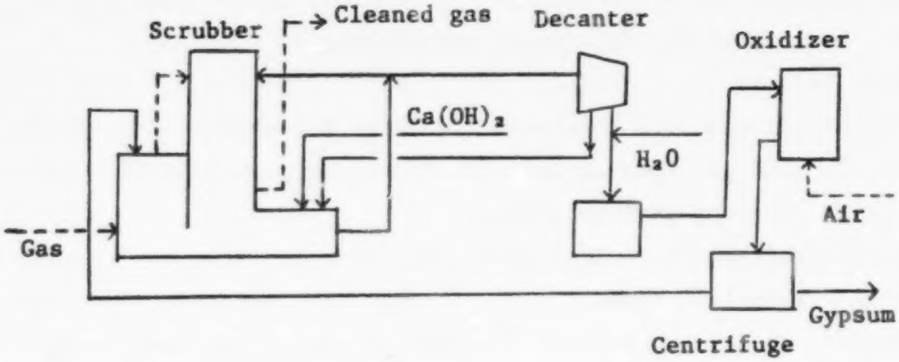

Figure 1 Kobe Steel $CaCl_2$-CaO/Gypsum process

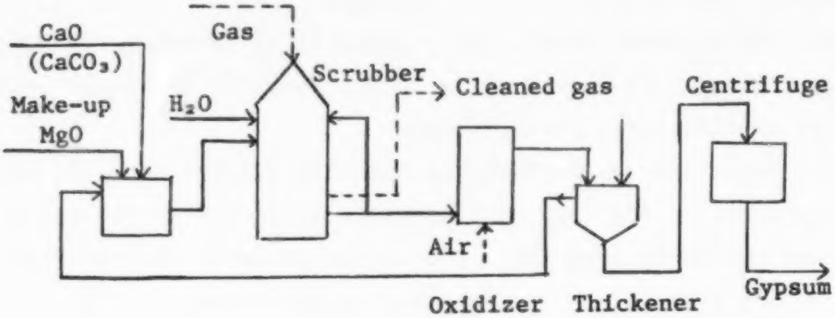

Figure 2 Kawasaki Magnesium-Lime/Gypsum process

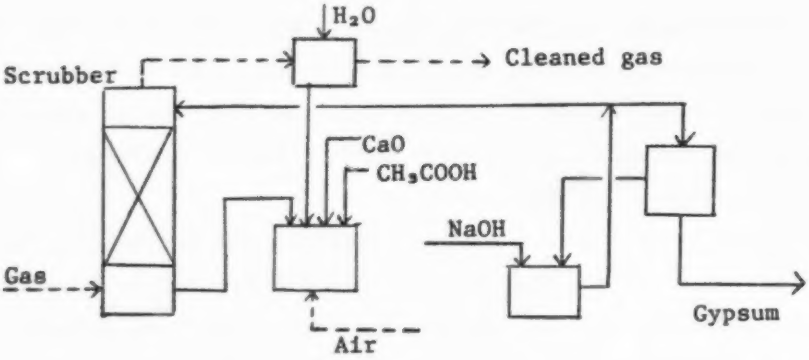

Figure 3 Kureha Sodium Acetate-Lime/Gypsum process

Two plants are in operation treating flue gas from an industrial boiler. In the early period of operation, soft scale formed in the piping of the scrubber system and in the mist eliminator. The problem was solved later and the plants have since been run at over 99% operability. Although MgO tends to inhibit the crystal growth of gypsum, the by-product gypsum is useful for cement and wallboard.

2.3 Nippon Kokan Process

Nippon Kokan (NKK) recently completed two large FGD plants to by-produce ammonium sulfate from SO_2 in flue gas from an iron-ore sintering plant and NH_3 in a coke oven gas. Details of the process have been reported1).

Some operation data of the Keihin plant are shown in Table 4. Flue gas (1,120,000 Nm^3/hr, 380 MW equivalent) is treated in two screen type scrubbers installed in parallel. A NH_4HSO_4 liquor leaving the scrubber is sent to an ammonia absorber where the gas is contacted with the coke oven gas containing NH_3 to form $(NH_4)_2SO_3$. Most of the $(NH_4)_2SO_3$ liquor is returned to the SO_2 absorbing system while the rest is oxidized by air to produce $(NH_4)_2SO_4$. The flue gas leaving the scrubber is passed through a wet electrostatic precipitator to eliminate plume.

Another plant, the Fukuyama plant, has a capacity of treating 760,000 Nm^3/hr of flue gas with one scrubber. The gas leaving the scrubber is heated to reduce plume without using a wet electrostatic precipitator.

The process is advantageous in producing ammonium sulfate from SO_2 and NH_3 in waste gases. H_2S present in the coke oven gas is removed before the gas is sent to the absorber because H_2S would form $(NH_4)_2S_2O_3$ which prevents the oxidation of $(NH_4)_2SO_3$ to the sulfate. Both plants have been operated trouble-free.

2.4 Dowa Aluminum Sulfate-Limestone Process

Three plants based on the Dowa process (Dowa and Naikai) have been operated smoothly. Although pressure drop of gas through the scrubber increased by about 50% in one year, plant operation has not been hindered. The scrubber packing is cleaned once a year during a scheduled shutdown of the gas source. Two new plants went into operation in late 1976. Three plants are under construction, one of them in China. The capacity of those new plants ranges from 30,000 to 140,000 Nm^3/hr.

2.5 Kureha Sodium Acetate-Lime Process

Kureha has continued pilot plant tests and improved the process2). By substituting lime for limestone, the size of the scrubber and the reactor as well as the evaporation of acetic acid from the SO_2 absorbing section has been substantially reduced. No acetic acid recovery section is needed because the acid is sufficiently recovered in a mist eliminator. Air is blown into the reaction tank eliminating a separate oxidation tower. The overall FGD cost seems to be lowered considerably even though lime is used. The process is applicable to simultaneous NO_x and SO_2 removal by adding a few vessels (3.3.3).

2.6 Chiyoda Limestone/Gypsum Sparger Scrubbing Process

Chiyoda has developed a new process to by-produce gypsum using limestone. The process will be described by another author in the present symposium.

3. SIMULTANEOUS REMOVAL OF SO_2 AND NO_x

3.1 Problems of Combination of FGD and NO_x Removal by Selective Reduction

NO_x removal from flue gas is increasing in importance with the growing consumption of fossil fuels, particularly coal. Selective catalytic reduction (SCR) to convert NO_x to N_2 and H_2O by reaction with ammonia using a catalyst has been considered most promising in Japan and applied to about 40 commercial plants. In those plants 85-95% of NO_x in flue gas is removed at 300-400°C with an NH_3/NO_x mole ratio of 1.0-1.2 and a space velocity of 5,000-10,000 hr^{-1}. Plant operation is easy with a clean gas but not with dirty gases containing dust and SO_x because they involve the following problems: (1) Plugging of catalyst with dust. (2) Poisoning of catalyst by SO_x, especially by SO_3. (3) Formation of NH_4HSO_4 in the air heater resulting in corrosion and plugging. (4) Consumption of large amounts of ammonia. (5) Large energy requirement for gas heating when SCR is applied after wet FGD. (6) Accumulation of NH_3 in the FGD scrubber liquor when SCR is applied ahead of FGD. (7) Some of the catalysts oxidize a portion of SO_2 to SO_3.

In the early stage of development, wet type FGD was applied ahead of SCR, as in No. 3 of Figure 4, in order to clean the gas to avoid problems (1) and (2) above. The combination, however, requires much energy for heating the gas after FGD. Yet, a considerable portion of SO_3 which is not caught by FGD causes problems (2) and (3) above. Moreover, mist from FGD aggravates the problems. Use of a high efficiency wet electrostatic precipitator after FGD (ahead of a heat exchanger) can solve the problems as has been done at the Chiba plant, Kawasaki Steel, which treats 800,000 Nm^3/hr flue gas from an iron-ore sintering plant. But such a system is very expensive and may not be applicable to other plants.

A recent trend in Japan is to treat gas at 350-400°C from a boiler economizer by SCR and then by FGD (No. 4, Figure 4). Catalysts based on TiO_2 resistant to SO_x have been produced. Present efforts are concentrated on developing reactors and catalysts free from dust plugging. Use of parallel flow by parallel passage reactor or a honeycomb type, tube type or plate type catalyst seems promising. Still problems (3) and (4) above remain unsolved and (6) is added. About 1% heat loss may be caused by problem (3) in addition to about 1% for reheating after FGD.

Selective non-catalytic reduction (SNR) which uses 1-2 moles NH_3 to 1 mole NO_x at 950-1,000°C has been tested in several large-scale plants (Table 5). In those plants, 40-50% of NO_x has been removed at a relatively low cost but problems (3), (4) and (6) seem larger than for SCR (No. 5, Figure 4).

Both systems No. 4 and No. 5 require a device to maintain an optimum reaction temperature and cope with the fluctuation of boiler load.

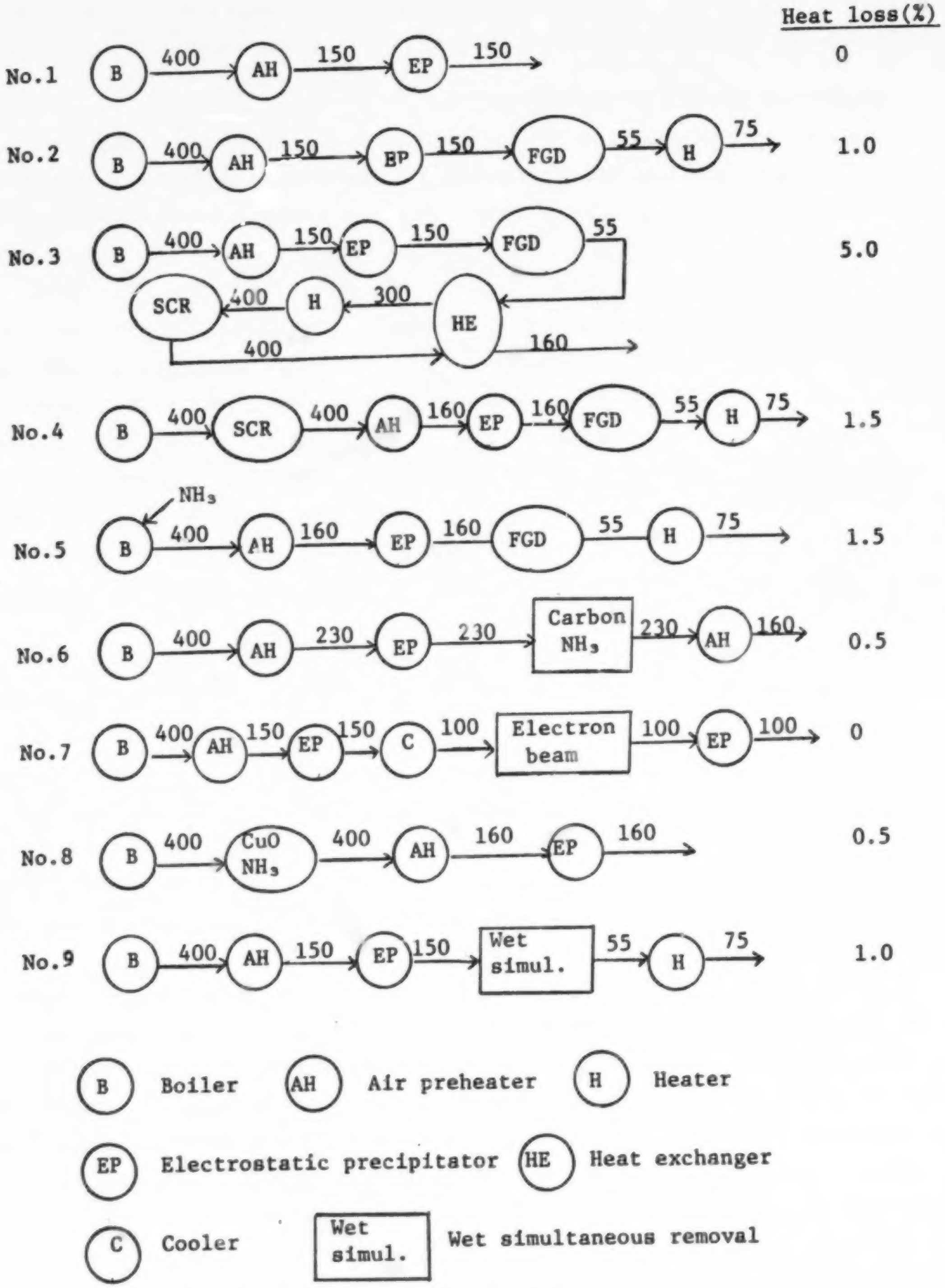
Heat loss(%)
No.1 B 400 AH 150 EP 150 0
No.2 B 400 AH 150 EP 150 FGD 55 H 75 1.0
No.3 B 400 AH 150 EP 150 FGD 55 HE 300 H 400 SCR 400 HE 160 5.0
No.4 B 400 SCR 400 AH 160 EP 160 FGD 55 H 75 1.5
No.5 NH₃ B 400 AH 160 EP 160 FGD 55 H 75 1.5
No.6 B 400 AH 230 EP 230 Carbon NH₃ 230 AH 160 0.5
No.7 B 400 AH 150 EP 150 C 100 Electron beam 100 EP 100 0
No.8 B 400 CuO NH₃ 400 AH 160 EP 160 0.5
No.9 B 400 AH 150 EP 150 Wet simul. 55 H 75 1.0
B Boiler
AH Air preheater
H Heater
EP Electrostatic precipitator
HE Heat exchanger
C Cooler
Wet simul. Wet simultaneous removal

Figure 4 Combined and simultaneous removal systems (Numbers show temperatures, °C.)

To avoid the problems of combination of wet FGD and dry NO_x removal, simultaneous SO_x and NO_x removal processes, either dry or wet, have been tested in pilot plants, and applied to a few small commercial plants.

Table 5. MAJOR PLANTS BY NON-CATALYTIC AMMONIA REDUCTION
(ALL TREAT FLUE GAS FROM OIL-FIRED BOILERS)

Plant constructor	Plant owner	Plant site	Amount of gas treated (Nm^3/hr)	Year of completion
Tonen Technology	Tonen Petrochemical	Kawasaki	540,000	1975
Tonen Technology	Mitsui Petrochemical	Chiba	200,000	1976
Mitsubishi H.I.	Mitsubishi Chemical I.	Mizushima	468,000	1975
Mitsubishi H.I.	Chubu Electric	Chita	1,036,000	1977

3.2 Dry Simultaneous Removal Processes

3.2.1 Activated Carbon Process. Activated carbon containing a small amount of heavy metal catalyst has been used in three pilot plants (4,500 Nm^3/hr plant of Unitika and two smaller ones) for simultaneous SO_2 absorption and NO_x reduction by ammonia. Lower temperature is favored for the SO_x absorption while higher temperature is suitable for the NO_x removal (Figure 5). Actually the plants are operated at about 230°C with a SV of around 1,000 hr^{-1} removing about 90% each of SO_x and NO_x. SO_2 absorbed by the carbon forms H_2SO_4 and NH_4HSO_4. After the reaction, the carbon is heated to 400°C in a reducing gas to convert the H_2SO_4 and NH_4HSO_4 to N_2 and concentrated SO_2. Regeneration by water wash is not suitable because the carbon is useless for NO_x removal when wet.

Ammonia consumption increases not only with NO_x but also with SO_x concentrations, while carbon consumption increases with SO_2. The SV value is from 1/5 to 1/10 of that for SCR, meaning that a large amount of carbon and a large reactor are needed. Therefore, the process may be useful in treating a relatively small amount of gas with a relatively low concentration of SO_x.

3.2.2 Electron Beam Radiation Process. A flue gas at about 100°C to which a small amount of ammonia is added is exposed to electron beam radiation. A powdery product consisting of ammonium nitrate and sulfate is formed and is caught by an electrostatic precipitator. Over 90% of NO_x is removed with 70-80% of SO_2 (Figure 6). The process has been developed by Ebara Manufacturing Co. jointly with the Japan Atomic Energy Research Institute and is going to be tested by Nippon Steel in a pilot plant with a capacity of treating 3,000 Nm^3/hr flue gas. Energy consumption is claimed to be not more than in other simultaneous removal processes but the investment cost seems high. Many large electron beam generators are needed for a large-scale plant. The process may suit gases relatively rich in NO_x and low in SO_2 and the by-product may be useful for fertilizer.

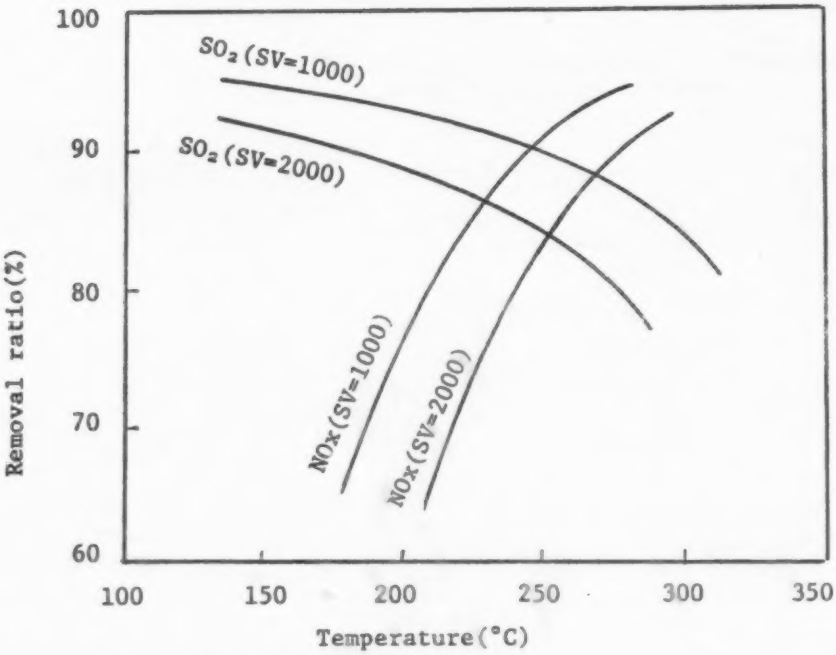

Figure 5 Efficiency of simultaneous removal by activated carbon and ammonia

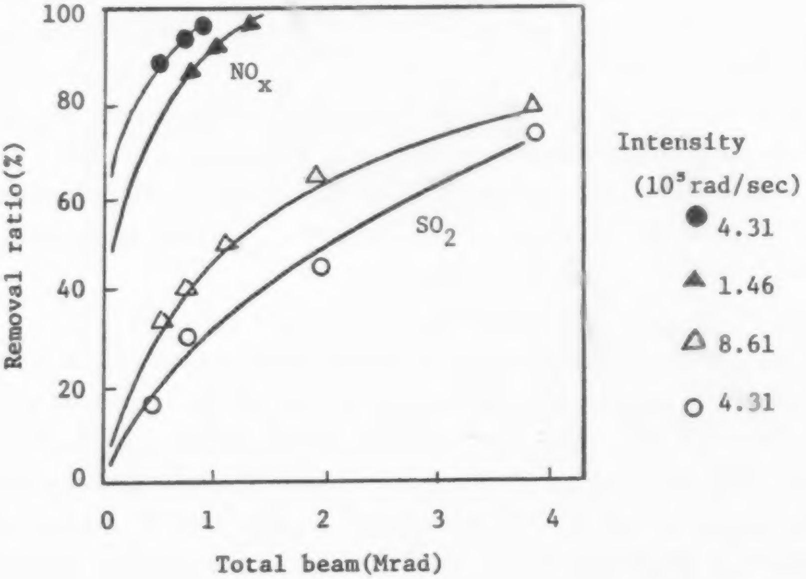

Figure 6 Simultaneous removal by electron beam radiation[3)]

3.2.3 Shell Copper Oxide Process. This process uses copper oxide as the SCR catalyst. Granular alumina impregnated with copper oxide serves as the absorbent of SO_2. The Yokkaichi plant, SYS, with a capacity of treating 120,000 Nm^3/hr flue gas from an oil-fired boiler by the Shell process, has introduced ammonia into a reactor at 400°C since 1975. Copper sulfate formed by SO_2 absorption as well as copper oxide works well but metallic copper formed by hydrogen reduction of the sulfate is useless. The overall NO_x removal efficiency is described as 65% with 0.99 mole NH_3 per mole of NO4).

3.3 Wet Simultaneous Removal Processes

3.3.1 Outline. For NO_x removal by wet processes, sodium scrubbing and magnesium scrubbing have been applied in Japan in pilot and small commercial plants, but those are not suitable for large plants because of the by-production of sodium and magnesium nitrates which have little use.

For large-scale application, NO_x should be converted to N_2 or NH_3. This can be achieved by utilizing the reducing effect of SO_2 in flue gas. Processes based on this principle are shown in Table 6. In the oxidation reduction process, NO which is fairly inactive is first oxidized to NO_2 and is absorbed in a limestone slurry containing a small amount of catalyst, while in the reduction process NO is absorbed in a solution of sodium or ammonium compound containing ferrous sulfate and a chelating compound such as EDTA (ethylenediamine tetraacetic acid) which forms an adduct with NO to promote the absorption.

Table 6. MAJOR WET SIMULTANEOUS REMOVAL PROCESSES

Process developer	Oxidizing (Complex) agent	Absorbent (Precipitant)	By-product	Plant capacity (Nm^3/hr)
Oxidation reduction process				
Mitsubishi H.I.	O_3	$CaCO_3$	Gypsum, NH_3	2,000
Ishikawajima H.I.	O_3	$CaCO_3$	Gypsum, N_2	5,000
Sumitomo-Fujikasui	ClO_2	$CaCO_3$	Gypsum, nitrate N_2, chloride	500
Reduction process				
Chisso Engineering	EDTA, Fe^{2-}	NH_3	$(NH_4)_2SO_4$	500
Kureha Chemical	EDTA, Fe^{2-}	CH_3COONa CaO	Gypsum, NH_3, (N_2)	5,000
Asahi Chemical	EDTA, Fe^{2-}	NaOH $CaCO_3$	Gypsum, N_2	600

Various reactions occur in the scrubber liquor. Most of the absorbed NO_x forms an imidodisulfonate $NH(SO_3M)_2$ or a sulfamate NH_2SO_3M (M = NH_4, Na, or 1/2Ca) by reactions (1), (2) and (3).

$$2NO + 5SO_2 + 6MOH = 2NH(SO_3M)_2 + M_2SO_4 + 2H_2O \tag{1}$$

$$2NO_2 + 7SO_2 + 10MOH = 2NH(SO_3M)_2 + 3M_2SO_4 + 4H_2O \tag{2}$$

$$NH(SO_3M)_2 + H_2O = NH_2SO_3M + MHSO_4 \tag{3}$$

Those equations indicate that 2.5 or 3.5 moles of SO_2 is needed to each mole of NO or NO_2. Since a portion of SO_2 is oxidized by O_2 in the liquor, 3.5-4.5 moles of SO_2 is required to attain about 90% NO_x removal.

The imidodisulfonate and sulfamate are decomposed in different ways depending on the process. In the Mitsubishi, Chisso, and Kureha processes, NH_3 or $(NH_4)_2SO_4$ is produced by reactions (3) to (6), while in the Asahi and Ishikawajima processes, N_2 is formed by reactions (7) and (8).

$$NH_2SO_3M + MHSO_4 + H_2O = NH_4HSO_4 + M_2SO_4 \tag{4}$$

$$NH_4HSO_4 + NH_3 = (NH_4)_2SO_4 \tag{5}$$

$$NH_4HSO_4 + CaO = CaSO_4 + NH_3 + H_2O \tag{6}$$

$$2NH(SO_3M)_2 + 1/2O_2 \xrightarrow{\text{heat}} 2M_2SO_4 + 2SO_2 + N_2 + H_2O \tag{7}$$

$$2NH_2SO_3M + 2M(OH)_2 + O_2 \xrightarrow{\text{heat}} N_2 + 2M_2SO_4 + 4H_2O \tag{8}$$

3.3.2 Oxidation Reduction Processes. A simplified flowsheet of the Mitsubishi (MHI) and Ishikawajima (IHI) processes is shown in Figure 7. Flue gas to which ozone is added is treated in a system similar to the wet FGD system by a limestone slurry containing catalysts (small amounts of NaCl and $CuCl_2$ in the IHI process)$^{5)}$. By using 1 mole ozone to 1 mole NO, about 90% of NO_x is removed together with over 95% of SO_2. An oxidation step of calcium sulfite is virtually unneeded because of the oxidizing effect of NO_2. A portion of the liquor from a gypsum centrifuge containing calcium imidodisulfonate and sulfamate, which are both water-soluble, is treated in different ways. In the MHI process, the compounds are hydralized to ammonium bisulfate, which is treated with lime to produce NH_3 and gypsum (reactions (3), (4) and (6)), while by the IHI process the liquor is evaporated and the solids are calcined to produce calcium sulfate and N_2 (Reaction 8). IHI also has a process to by-produce NH_3 in a way similar to that of MHI.

The overall reactions of the two processes may be simply expressed by equations (9) and (10).

$$2NO_2 + 7SO_2 + 7CaO + 3H_2O = 7CaSO_4 + 2NH_3 \tag{9}$$

$$2NO_2 + 5SO_2 + 5CaO + 1/2O_2 = 5CaSO_4 + N_2 \tag{10}$$

The use of ozone equimolecular to NO to achieve 90% removal results in the formation of a small amount of nitrate which would accumulate in the liquor. For 80% NO_x removal, less ozone is used and the nitrate formation is avoided.

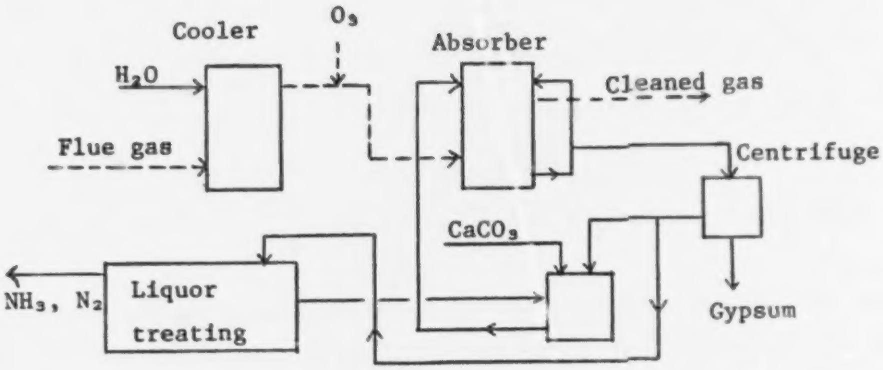

Figure 7 Simplified flowsheet of wet-limestone simultaneous removal process (oxidation reduction process)

A great advantage of the processes is that they can be carried out in conventional wet lime/limestone scrubbing plants by adding a small liquor treatment system. On the other hand, a major drawback is the consumption of ozone. For flue gas containing more than 200 ppm NO, ozone oxidation is too expensive. The ammonia by-production process may be useful for flue gas from coal if NO content is reduced to 250 ppm by combustion modification and to 150 ppm by injecting the by-product NH_3 into the boiler.

Sumitomo Metal jointly with Fujikasui has developed a process that uses ClO_2, which is less expensive than ozone. By that process, about a half of NO_x in the gas is converted to N_2 while the rest goes into a scrubber liquor as calcium nitrate. In addition, the liquor contains a considerable amount of calcium chloride. The process may be useful if the by-product liquor, a mixture of calcium chloride and nitrate, can be utilized. One possible use is as an anti-freeze agent for roads. Otherwise, the liquor may be treated with ammonium sulfate to precipitate gypsum and to recover ammonium chloride and nitrate liquors which may be used as fertilizer.

3.3.3 Reduction Processes. Reduction processes (Chisso, Kureha and Asahi) use a sulfite solution containing ferrous ion and EDTA, which promote the absorption of NO, forming an adduct. EDTA is fairly expensive (about $3,000/metric ton) but is consumed in a very small amount. A small portion of the absorbed NO may be converted to N_2 by reaction (11) but a major portion forms an imidodisulfonate. A large portion of absorbed SO_2 forms a dithionate, $(NH_4)_2S_2O_6$ or $Na_2S_2O_6$, by the effect of ferrous ion and EDTA. Therefore, a decomposition step of those compounds is needed in addition to a standard FGD process.

The Chisso process uses ammonia scrubbing while the Kureha process uses a sodium acetate-lime double alkali system. In those processes imidodisulfonate and dithionate undergo hydrolysis at 120-140°C under pressure to form a bisulfate and sulfate by reactions (3), (4) and (12). The overall reactions of the two processes may be expressed by equations (14) and (15), respectively.

$\mathrm{Fe - EDTA{\cdot}NO + SO_3^{\,2-} \longrightarrow Fe - EDTA + SO_4^{\,2-} + 1/2N_2}$ ....(11)

$\mathrm{M_2S_2O_6 + H_2O + 1/2O_2 = 2MHSO_4}$ ....(12)

$\mathrm{M_2S_2O_6 \xrightarrow{heat} M_2SO_4 + SO_2}$ ....(13)

$\mathrm{2NO + 5SO_2 + 8NH_3 + 8H_2O = 5(NH_4)_2SO_4}$ ....(14)

$\mathrm{2NO + 5SO_2 + 5CaO + 3H_2O = 5CasO_4 + 2NH_3}$ ....(15)

The Asahi process is a combination of a sodium-limestone FGD process and thermal decomposition of the imidodisulfonate and dithionate (reactions (6) and (11)). SO_2 released by the decomposition is used to decompose Na_2SO_4 to $NaHSO_3$ and $CaSO_4$ in the same manner as in the above FGD process. The by-products of the Asahi process is gypsum and N_2 with a small amount of Na_2SO_4.

Those reduction processes do not require any oxidizing agent and may suit NO_x-rich gasses provided that more than about 4 moles of SO_2 is contained for each mole of NO. However, since the absorption of NO occurs slowly and its solubility is small, the scrubber requires many stages with a heavy pressure drop and a large L/G ratio. Although 90% NO_x removal can be attained, 70-80% removal may be a more practical target.

3.4 Summary

Relationships of SO_x and NO_x concentrations in gas to suitable processes for treatment is illustrated in Figure 8.

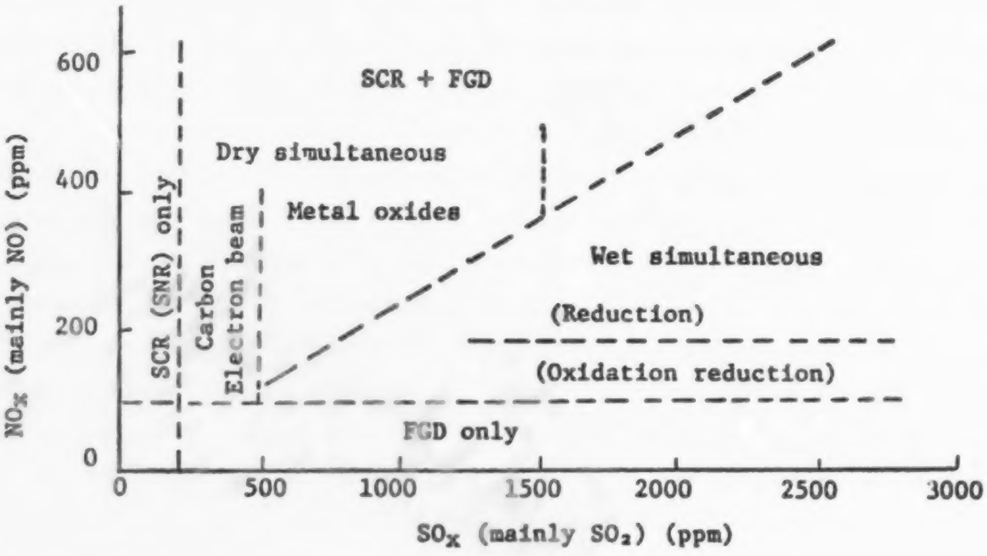

Figure 8 Gas composition and suitable processes

SCR and SNR are suitable for gases with low SO_x concentrations. Although they can be used for SO_x-rich gases in combination with FGD, ammonium bisulfate formation will present a problem. Dry simultaneous carbon and electron beam processes would suit gases with relatively low SO_x concentrations. For the Shell process, a high SO_x concentration will necessitate a frequent switching between absorption and regeneration which is not desirable in obtaining a high NO_x removal efficiency.

On the other hand, SO_2-rich gas would suit wet simultaneous removal processes because NO_x removal efficiency increases with SO_2 concentration. Oxidation reduction processes would fit gases with relatively low NO concentrations.

Developers of wet simultaneous removal processes claim both investment and operation costs for processes for flue gas from an oil-fired boiler are about 40% more than those for FGD alone and slightly less than those for a combination of SCR and FGD. The costs for wet as well as dry simultaneous removal processes are uncertain yet. Further improvements are desired for any combination or simultaneous processes for wide application on a large scale.

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DESULFURIZATION TECHNOLOGY IN THE FEDERAL REPUBLIC OF GERMANY

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ABSTRACT

This paper presents an overview of the standards and regulations in effect in the Federal Republic of Germany (FRG) for SO_2 control. Also discussed from both a technical and economic point of view are the current status and plans for flue gas desulfurization systems in the FGR.

Desulfurization Technology in the Federal Republic of Germany

Coal is the only energy.resource available in the Federal Republic in more than nominal quantities. The desired higher independency of imported oil and natural gas postulated in the energy program of the Federal Republic can thus be only obtained if coal is used to a larger extent to meet the power demand than has been done up to now. Development work required to arrive at this end is performed within the context of the Energy Research Program of the Federal Government.

Two obstacles are obstructing the increase of a consumption of hard coal for the generation of electrical power:

1) Coal is very expensive (22 DM/Gcal, 2.5 \$/MBTU).

2) Emission limitation standards are stringent already today. A further drastical tightening of the laws is to be expected. This is true in particular for sulfur emissions.

According to the rules valid up to summer 1977 a limit of 3.75 g SO_2/kWh_{el} was not allowed to be exceeded. Because of the uncertainty of the technical availability of desulfurization equipment even a stepwise implementation of this regulation was conceeded. Only after a test period of 3.5 years during which the desulfurization equipment was allowed to run to merely 40 pct. of its capacity the above mentioned value became mandatory. It was also left to the operating company whether this value was obtained by treatment of a large bypass stream with low efficiency or by nearly complete desulfurization of a small fraction of the flue gas.

Since 1977 a much stricter rule is enforced in Northrhine-Westphalia, the state in the Federal Republic extracting most of the coal. The maximum sulfur dioxide emission must not be higher than 2.75 g SO_2/kWh_{el}. As an additional aggravation it was postulated that the total of 100 pct. of the flue gas has to be treated. This additional postulation guarantees that other pollutants in the flue gas as fluorine, chlorine, and dust are removed to a very high degree. The amount of NO_x is also reduced considerably.

While this rule is about to be enforced a further drastical lowering of the limits is being discussed. The Federal Authorities are postulating a value of 1.25 g SO_2/kWh_{el}. By some power plant operators this limit is considered to be prohibitive for the use of coal for electric power generation. This value corresponds to 0.3 lb SO_2/MBTU by the definition used in the USA. Figure 1 shows which degree of desulfurization has to be achieved in order to obtain the various values mentioned with coal of varying sulfur content. Our utilities would be happy if the previous limit--fiercely opposed in those days--would be still the standard today. And this value is still lower than the standards presently valid in the USA.

The stricter standards are on the other hand a strong stimulus to develop more effective and more economical desulfurization methods and to increase the efforts to develop more advanced electricity generation concepts.

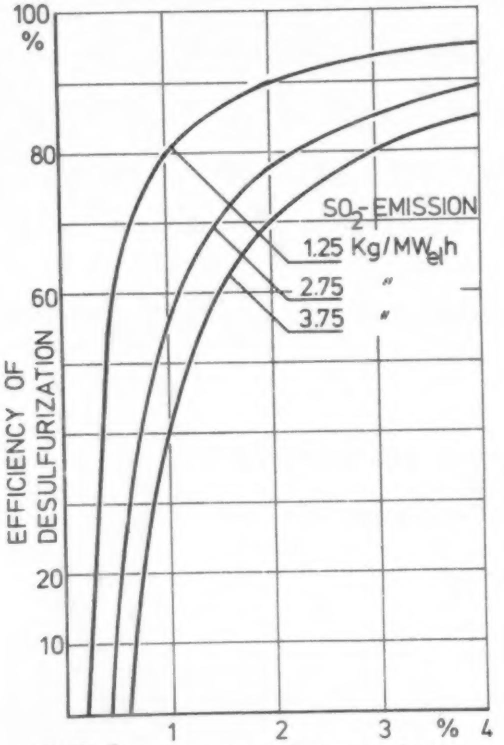

FIG.1 SULFUR CONTENT IN COAL

From the point of view of the Federal Republic it is also the high price of coal which demands the application of new processes. All processes under consideration promise a markedly higher efficiency wherefore fuel costs lose their significance for the costs of electricity generation. Due to the fact that in Germany the allowable SO_2 emission is based on the quantity power generated it is also simpler to meet the given standards. All new processes being under development here are characterized by the fact that the desulfurization is an integral part of the process and not an appendage as in case of the flue gas desulfurization. This results in markedly lower investment and operating costs for the observance of the same emission rates.

The following advanced processes are being developed in the FRG (Figure 2):

Atmospheric Fluid Bed Combustion (AFB)
Pressurized Fluid Bed Combustion PFB
Low-BTU-Combined Cycle
VEW-Process

All these processes--except for the VEW-Process--are also extensively being worked at in the USA.

To our knowledge the following state of affairs can be recognized in this "competition" presently:

Fig.2 Advanced Power Generation Processes
(R+D-Program of the FRG)

	power-plant efficiency	1.25kg/MW_{el}h achievable (1.5% sulfur in coal)	technically available
atmospheric fluidized bed	37 %	(yes)	1982
pressurized fluidized bed	41 %	yes	1985
Low-BTU-combined cycle	43 %	yes	1985
VEW-Process	41 %	no	1983

AFB: The US do have the lead as it is proven by the 30 MW plant in Rivesville. A plant of the same capacity is under design in Germany.

PFB: Both the US and the FRG have reached the same level. Both countries cooperate in the IEA experimental plant in England (85 MW) and are constructing in addition domestic experimental plants (20 MW) in order to evaluate the application possibilities of gas turbines.

Low-BTU-Combined Cycle: The FRG does have a significant head start as is proven by construction and operation of a 170 MW plant. Even if present results may be not satisfactory and later power plants of this type may have to be constructed differently the experience accumulated so far will be useful to full extent. A lead derives also from the fact that the gasification processes under development in the Federal Republic do better fit into power plants. In this context one should mention the statement of intent signed by your Secretary of Energy and our Minister of Research and Technology which says that a common development of a Low-BTU-gasifier will be undertaken in Germany.

Only experts will probably be familiar with the VEW-Process. I should like to give a short description of the process as it has some advantages which could be of importance for the USA as well. (Figure 3)

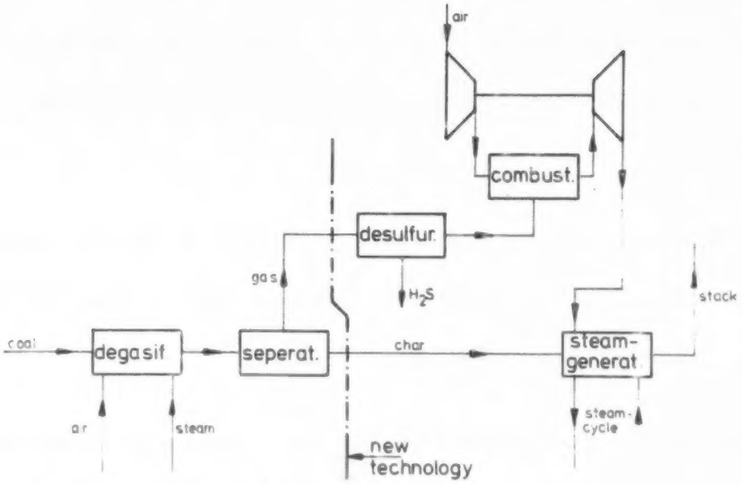

Fig. 3 VEW-PROCESS

Powderized coal is heated to 1000°C (1768°F) by a partial combustion. It is thus degassed and subsequently gasified to some extent. At this step a large fraction of the sulfur is removed. After separation of the coke the gas is desulfurized and used in a gas turbine. The exhaust gases of the turbine are burned in a steam power plant along with the partly desulfurized char.

We consider it a main advantage of this concept that only to a small extent a new technology is required and that these parts of the process do not pose any problems. The unconventional steps of the process have been tested in a pilot plant with a capacity of 1 ton coal per hour. The construction of a 15 t/h plant has just been started.

A disadvantage of the process is that only a 60 pct. desulfurization of the coal is obtained. If coal with a high sulfur content is used the SO_2 emission limits are exceeded.

We work on the development of the four mentioned processes with a high priority as they are suitable methods to convert coal with a high efficiency and little impact on the environment into electricity. We are, however, aware of the fact that these processes will be available technically only ten years from now and that the commercial use on a larger scale will not take place before 15 to 20 years. The challenge remains therefore to convert coal into electricity in the next future without detriment to the environment. Suitable means to this end are coal desulfurization and flue gas desulfurization. Because of the limited effectiveness of coal desulfurization, flue gas desulfurization is unavoidable.

If in the more advanced processes for the conversion of coal into electrical power something like an equal standard of the development in the USA and the FRG could be stated a clear lead of the USA in the flue gas desulfurization is evident. This is true, however, only for the commercial application and large scale proven technical viability, not necessarily for the future potential of the processes.

In the past three different processes have been developed in the Federal Republic which do fulfill all possible local demands. All three processes have reached a state where large scale technical application is possible.

The Bischoff-Process is being used already on a large technical scale. A 20 pct. flue gas fraction of a 720 MW power plant is being treated. The process uses a lime suspension in a two-step Venturi scrubber. The resulting sulfite sludge is compacted and placed on dumps. The plant went on stream in February 1977. So far there were some difficulties caused by crust formation in the heat exchanger used for reheating of the purified exhaust gas. The problem of incrustation have been solved by smaller plant modifications. So far the plant has been in operation for 1000 hours. Its efficiency according to specifications has been proven.

A process like the one just described can be resorted to in the Federal Republic only in exceptional cases as there are no possibilities for the deposition of the waste sludge due to the high population density. From the very beginning processes have therefore been emphasized which produce a re-usable material.

This postulation is most clearly fulfilled by the BF-Process. The process works as follows: The hot flue gases are conveyed through activated coke which absorbs the SO_2 physically as H_2SO_4. The desorption of the coke takes place in a 2nd step by heating to 600°C (1048°F). Subsequently the SO_2 which has a concentration of 30 pct. can be reworked into pure SO_2, H_2SO_4, or elementary sulfur. In a highly industrialized country there is a ready market for all these three products. Elementary sulfur can also be deposited without difficulties if this should be required.

The process was developed to technical maturity in a prototype plant which treats 150,000 Nm^3/h flue gas. This plant was put on stream two and a half years ago and has been in operation meanwhile for 7000 h. The availability is satisfactory, a removal efficiency of more than 80 pct. is obtained.

The advantages of this process are that a re-usable material is produced and that no reheating of the purified flue gas is required. It is certainly a question of the local conditions whether these advantages justify the higher investment costs.

The third process developed in the Federal Republic (Saarberg-Holter) does also fulfill the postulation for a re-usable product without being more complicated than conventional lime scrubbing. I may skip a technical description of the Saarberg-Holter-Process as on this subject another paper will be given during this symposium.

I would like to mention, however, the most essential aspects which brought us to the conclusion that this might be the appropriate process for the local conditions in the FRG.

- A re-usable final product is generated. The gypsum produced in this process is of a high purity and can be used in the construction industry, in cement manufacture, and in the mining industry.
- The process design is simple which results in low investment costs and an unusually high availability (95 pct.).
- Low power demand.
- High degree of desulfurization (95 pct.).
- Simple automatic control.

The results were so far obtained in a prototype plant with a gas through-put of 125,000 Nm^3/h. The operation time is 17,000 h in total. Since commercial plants will be built in modules no scale-up problems are foreseen.

An order for the partial desulfurization of one 700 MW plant by this method has so far been placed. In the context of plans for an enlargement of the coal-based power plant capacity in the Federal Republic one may expect further application possibilities in Germany.

In the spring of this year additional government funds have been made available in order to demonstrate new technologies for the lowering of emissions and to create thus the prerequisite for an increased utilization of coal. Flue gas desulfurization is in the center of these program. For the first time a plant will be constructed in the Federal Republic where the total of the exhaust gases of a 750 MW power plant will be desulfurized. In the context of a preliminary investigation to be completed in the current year suitable processes will be chosen. Two basic variants are being discussed.

1) A combination of a dry and a wet process.

2) A combination of two wet processes.

The advantage of the first concept is that a reheating of the flue gases can be omitted: At the dry desulfurization the purified gas has a temperature of 130°C (202°F). Mixing with flue gas desulfurized in a wet process renders a gas with a sufficiently high temperature.

An additional advantage lies in the fact that two different materials are produced which can be sold in a not too large surrounding area without glutting the market.

Presently, however, it is still uncertain whether these advantages will compensate for the probably higher investment costs and possible operating disadvantages.

In case of a decision exclusively for wet methods the reheating of the purified gases can be achieved by heat exchangers. To provide suitable heat exchangers an extensive development program has been initiated. Research is being done on materials, design, and cleaning possibilities.

Furthermore developments have been started to guarantee the marketability of the gypsum which is generated in large quantities by lime-based processes. The substitution of natural gypsum is intended as there is already a sufficient supply of synthetic gypsum in the Federal Republic. A mechanical pretreatment is required in order to enable the funneling of desulfurization gypsum into the normal processing ways of natural gypsum. Application experiments with the manufactured gypsum will conclude this development. Only when all side-products from a power plant have found a useful application the waste disposal problem can be considered to be solved.

Another project which has just been started seems to be worth to mention because it shows a new method of attack of the problem of exhaust gas treatment. The flue gases of a 200 MW power plant are subjected to wet scrubbing after a dedusting step. The desulfurization works according to the Saarberg-Holter-Process which allows a very compact design of the equipment. This compactness makes it possible to build the desulfurization equipment in the interior of the cooling tower. The stream of the cooling air is not impeded to an unacceptable extent.

From this combination we expect the following essential advantages:

- No smoke-stack is required for the flue gases. By mixing with the cooling air and the high linear momentum at the tower outlet a favorable distribution of the gases at all weather conditions is obtained.
- A reheating of the purified gases can be omitted.
- The desulfurization equipment is not conspicuous. The scenery is affected by the power plant only to a lesser degree.

This "model power plant" will go on stream in 1980 according to our present schedule. Maybe it will become the prototype of a new generation of power plants.

Summary:

The development of advanced power plant processes it promoted with emphasis in the Federal Republic. Three different methods have shown to have good prospects.

It is in the nineties at the earliest that the new processes will add to the power supply to a more than nominal extent. The application of flue gas desulfurization is therefore unavoidable in the meantime if electricity is to be generated from coal without detriment to the environment. There is a choice of three domestically developed processes in Germany.

Considerable additional funds have been provided by the government in the current year in order to invalidate the opposition against an increase of coal consumption for power generation. The funds will be used for the demonstration of new technologies which lower the emissions of power plants. We will be able to report on results concerning these plants only at one of the next symposiums.

EPRI'S FLUE GAS DESULFURIZATION PROGRAM, RESULTS, AND CURRENT WORK

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EPRI'S FLUE GAS DESULFURIZATION PROGRAM, RESULTS AND CURRENT WORK

EPRI has budgeted approximately 10 million dollars in R&D funds to Flue Gas Desulfurization (FGD) over the next five years. This paper reviews EPRI's approach to FGD, summarizes results and outlines ongoing projects.

Who is EPRI?

For those of you unfamiliar with EPRI, a brief introduction may be useful. EPRI is the research arm of the U.S. electric utility industry. We are a non-profit institute organized in 1972 to finance, coordinate, exchange and direct research work on behalf of the U.S. electric utilities. We are not an advocate organization and represent both public and private utility interests. Eighty percent of U.S. electrical generation capacity is operated by member organizations. Member organizations are assessed a fee (in 1976 it was .14 mills/kWh) as a means of program funding. We fund many R&D areas including generation, transmission and utilization of electricity by environmentally acceptable means.

What are We Doing?

EPRI's Air Quality Control Group is reporting state-of-the-art information, evaluating designs and process subsystems, testing improvements and sponsoring process development leading to a reliable, site-specific design basis for utility use of FGD. EPRI's emphasis in SO_x control is shifting from a strict reporting role to a development, evaluation and reporting role. This reflects the utility commitment to FGD technology, the prospect of tighter SO_x emission standards and the realization that FGD is the least expensive route to the required levels of positive control of sulfur dioxide emissions from new coal-fired boilers. While low-sulfur fuel options and tall stacks for polltant dispersion continue to be pressed by the utilities as lower cost SO_2 control options, the pressure for positive control is increasing. If proposed new source performance standards (NSPS) are promulgated at the proposed 0.4-0.8 lbs SO_2/million Btu level, and/or for 90% control, the low-sulfur coal and tall stack options will be virtually eliminated. The utility commitment is already on the order of 50,000 MW and $5 billion to FGD technology. The need is for more efficient, reliable and environmentally accceptable processes than can satisfy the utility industry's increasing demand for coal-fired generation. The cost and time to develop alternate coal conversion options was underestimated a few years ago, and it appeared possible to go to these options without a large FGD development effort. Now utilities are actively seeking FGD systems with improved reliability and lower cost. EPRI's program will initially emphasize providing site-specific design and operating bases that allow successful scrubber performance, operability and reliability. Longer-term emphasis is on (1) supporting the development and demonstration of improved FGD systems that minimize byproduct disposal water pollution, and land use impact; and (2) maintaining data on the state-of-the-art and associated cost of SO_x control technology so that the utility industry can evaluate the cost-benefit trade-offs of various levels of control. Whenever possible, EPRI will monitor utility-sponsored demonstration projects and participate in testing new processes, making useful data available to the industry.

EPRI plans to characterize individual full-scale lime/limestone scrubbing units that are selected to represent state-of-the-art FGD installations. These units will be fully measured and characterized on all liquid, gas and solid streams. Engineering designs, decision and compromises will be reviewed and evaluated. All aspects of the scrubber will be reviewed as well as material handling, alkali preparation, water treatment and sludge disposal.

Process innovation and design has historically been the responsibility of the system supplier. Now utilities and EPRI are attempting to increase reliability and lower cost through establishing a site-specific design for FGD. EPRI's approach is to investigate equipment components and provide individual utilities with enough information to put the best pieces together and create a reliable system. EPRI is now undertaking non-regenerable FGD development work on promising systems at the pilot plant scale (on TVA's Colbert Station Pilot Plant), a prototype demonstration scale (at TVA's Shawnee Station), and in theoretical modelling of lime/limestone chemistry. Regenerable FGD has been evaluated and specific subloops with promise are being developed under EPRI sponsorship.

Program Objectives

Specific objectives to be achieved are:

- By 1979, establish and operate prototype-scale scrubber development and evaluation centers for both eastern and western coal.
- By the final quarter of 1978, develop the design and operating support basis that will allow the utility industry to confidently purchase, if necessary, large-scale, closed-loop, lime/limestone scrubber and associated byproduct disposal systems for high-sulfur coal combustion.
- By 1980 complete the pilot-scale (20-50 MW) development, tests, and evaluation of advanced regenerable stack gas desulfurization processes and subloops on coal-fired utilities.
- By 1980 demonstrate systems to retrofit lime/limestone designs to eliminate calcium sulfite sludge and produce salable byproducts.
- By 1980 demonstrate subsystems to produce elemental sulfur without the use of reducing gas.
- By 1980 develop method to regenerate spent sodium/sulfur compounds from wet and dry FGD processes.

The EPRI program in SO_x control is generally as shown in the attached bar chart (Table 1). This planning document outlines the SO_x program and funding plans through 1982. This bar chart is presently being updated to reflect small changes in EPRI's program. The details of demonstration work in later years will be clearer as early evaluation and design projects are completed. Results are highlighted below, with more details in the project sections.

TABLE 1

SOx CONTROL PROGRAM Thousands of Dollars

OBJECTIVE	PROJECT DESCRIPTION	CONTRACTOR	RP NO.	1975	1976	1977	1978	1979	1980	1981	1982
Develop site specific design basis for lime/limestone scrubbing	Evaluate utility operating experience	Battelle	209	150							
	Unit studies-demister, reheat, cost variation	Battelle	209-1		175						
	1MW tests-reheat, corrosion, advance scrubber config.	TVA	537		633						
	Evaluation of F.G.D. process control capability	So. Cal Edison Sterns-Roger	630		200						
	State-of-the-art design guidelines for lime/limestone scrubbing	PEDCO	982-1			100	200 ▲				
	Low sulfur/alkaline ash scrubbing characterization	A.D. Little	785			100					
	Chemistry modifications to improve reliability & cost	RFP	1031			200	500	200 ▲			
	Commercial system dev. & technical field support (eastern high sulfur coal) SHAWNEE	TVA	1033			1800	1500	500	500	500	500
	Commercial system dev. & technical field support (western low sulfur coal)		Source Sought					200	500	500	
	Evaluation of sludge dewatering processes	Envirotech	786-2			150	200 ▲				
	Sludge composition & leachability	So. Cal Edison Radian	202			200					
	Sludge composition	Radian & M. Baker	786-1			150					
Develop viable regenerable F.G.D. technology	Comparative evaluation of regenerable F.G.D. processes	Radian	535	150							
	Comparative design of advanced processes	Stone & Webster	784		500	1000					
	Aqueous carbonate process development (Niagara Mohawk)	Atomics International	981			150					
	Co-sponsored pilot construction & operation of advanced processes (Regeneration sub-systems)						1250	2000	1000 ▲		
	Pilot test of chiyoda, double alkali, berghau processes	Southern Services	536		250	200					
	Demo of mag-ox process	Phila. Electric	1032			200					
	100MW + demonstration of advanced process(es)								600	1400	3500
Develop technical basis for intermittent control	Basis for intermittant control strategy	Econ. & Science Planning	50A 75-315	40							
Develop practical dry SO_2 removal technology	Technology evaluation of particulate & dry SO_2 removal	Bechtel	491	50							
	Feasibility of alternative reactants	Aerotherm	982 2			40					
	Laboratory process development					50	100				
	Pilot scale optimization (Arapahoe)	RFP						500			
	Commercial scale evaluation (Cameo)							100	200 ▲		
	Reactant regeneration						400	500	500		
			TOTAL	670	1800	4150	4250	4000	3300	2400	4000

RESULTS - LIME/LIMESTONE FGD

RP202 - Southern California Edison Company and Radian Corporation.

- Report - Discusses environmental effects of trace elements from ponded sludge and ash. Shows soil attenuation is significant. (EPRI 202, Final Report, September 1975.)

RP209 - Battelle Columbus Laboratories

- Report - Surveys SO_x scrubbing installations, shows flow sheets, statistics, photographs. (EPRI 209, Final Report, August 1975.)
- Report - discusses mist eliminator design, problems, factors to consider in specifications. (FP 327, Final Report, December 1976.)
- Report - Discusses reheat systems for SO_x scrubbers, energy requirements, possible configurations, design considerations. (FP 361, Final Report, February 1977.)

RP537 - TVA Colbert Test Facility

- Tests - Of horizontal (Weir) and cocurrent scrubbers.
- Tests - On reheaters, corrosion, and erosion of materials and of sludge correlations with operating conditions.
- Report - Now in draft form.

RESULTS - REGENERABLE FGD

RP535 - Radian Corporation

- Report - Compares 12 regenerable FGD processes showing flow sheets, chemical and energy balances, environmental impacts, material use; shows no single process is superior.

RP536 - Southern Company Services - 20 MW Tests

- Tests - This 20 MW demonstration program tested Chiyoda 101, CEA/ADL dual alkali and Bergbau-Forschung SO_2 removal processes.

Report - Now in draft form.

RP784-1 - Stone & Webster Engineering Corporation

- Evaluation - Reviewed design and cost of regenerable FGD to decide on best processes for detailed design.
- Report - Now in draft form.

RESULTS - DRY REMOVAL FGD

RP491 - Bechtel Corporation

- Report - Survey and assessment of dry SO_2 removal system status. Reviews theory and limits of backend sorbent injection for SO_2 removal. (FP 207, Final Report, October 1976.)

PROJECTS - LIME/LIMESTONE FGD

RP209 - Battelle Columbus Laboratories

Under this project, an initial assessment entitled, "Status of Stack Gas Control Technology," was published (EPRI 109, August 1975) and two subloop unit studies have been completed and reports published as noted below.

EPRI's Air Quality Control Program has funded a number of projects that will provide a firm data base for lime/limestone scrubber design. Battelle Columbus Laboratories evaluated the scrubbing installations across the U.S. (RP209). This project summarized the state of the art of lime/limestone scrubbing technology and highlighted a number of major generic problem areas that restricted the performance, reliability, and cost of this technology. To address these problems, the EPRI program has focused on each of the subsystems that make up a scrubber. These subsystems, which include reheaters, demisters, contactor-absorbers, recyle tanks, and dewatering devices must be designed properly to ensure their individual reliability and thus the reliability of the entire scrubber system. However, making the situation even more complicated, each subsystem must be designed to incorporate site-specific variations stemming from the high degree of chemical variability in coal. Thus, the success of a scrubber at one location does not ensure its success at another site.

Each scrubbing system must be a site-specific hybrid, incorporating subsystems; no single lime/limestone scrubbing system will satisfy the SO_2 emission requirements of the entire utility industry.

EPRI has recently completed unit studies on two subsystems identified in the Battelle-Columbus report as limiting lime/limestone scrubbing capability -- mist eliminators and reheaters.

An improperly designed mist eliminator builds mudlike solids and scale on mist eliminator blading. This plugs the assembly and requires the shutdown of the entire scrubbing system for cleaning. Good mist eliminator design results in lower operating costs by minimizing the amount of moisture carry-over that requires reheating along with the scrubbed flue gases. Excessive moisture carry-over can result in a significant waste of energy in the reheaters. However, in attempting to minimize the carry-over of moisture from the demister, the design engineer may design a demister subloop that plugs easily. This and other trade-off problems are addressed in the report, "Guidelines for the Design of Mist Eliminators of Lime/Limestone Scrubbing Systems," EPRI FP 327.

The report evaluates 15 parameters that must be considered in designing a reliable and efficient mist eliminator. It also summarizes the operational experience with mist eliminators of all scrubber systems presently operating in the U.S. The problems encountered with these operational mist eliminators and the design parameters that would have minimized or eliminated them are described.

After a flue gas is scrubbed, it is reheated from a saturated temperature of about 52°C (125°F) to a stack exit temperature of 80°C (175°F) to avoid downstream condensation, visible plume, and/or to disperse emissions. Reheating a scrubbed flue gas is the major energy requirement of a scrubbing system which amounts to 25-50% of the scrubbing system's energy needs. This can account for 1-3% of the total boiler input. Any savings in energy requirements for reheat will result in substantial decreases in operating costs of scrubbers. In a second unit study recently completed, "Stack Gas Reheat for Wet Flue Gas Desulfurization Systems," EPRI FP 361, guidelines are given on how to select the most efficient (energy input) reheater system for reheating a stack gas. The study evaluated reheat methods currently in use: in-line reheat, indirect hot air reheat, and direct combustion reheat. It was determined that the reheat temperature required depends on the objective for the reheat, while the minimum heat requirement depends on the degree of reheat, the method, and the quantity of mist carry-over from the scrubber.

The smallest amount of reheat is needed when only the avoidance of downstream condensation is required. (An in-line reheater requires the least energy.) The minimum calculated heat requirements for avoidance of downstream condensation showed that the indirect hot reheat and the direct combustion reheat required 12% and 18%, respectively, more heat input than that needed by in-line reheat.

As a final result of this project, a continuing stack gas emission coordination center was established. It continues to be maintained and updated by Battelle through EPRI and other industrial funding. This supplies detailed and coordinated information on currently operating SO_2 cleanup systems for the utility industry, so that the essential design and operating elements of reliable, effective SO_2 removal can be defined systematically and quickly. This informational base is continually updated by visits to specific utility stack control installations, hardware vendors, and technology developers to assemble the most current data which defines the factors influencing the performance, cost and reliability of stack gas control systems. The project funding allows the maintenance of a computer information storage and retrieval system so that utility personnel can receive quick responses to their specific pollution control inquiries. This service is available to utility personnel through EPRI.

RP537, "Development of Improved Lime/Limestone Scrubbing Technology" (Tennessee Valley Authority)

This project made use of the 1 MW pilot FGD plant facility of TVA Colbert Power Plant to expand the existing lime/limestone scrubbing technology data base.

This project has four tasks:

1. Study reheaters as a function of both technical and economic performance to establish a design basis. The reheat test program has, to date, measured the heat balance for two of the three reheat schemes included in the test program. The reheat test program also evaluated materials of construction and system operability. The in-line indirect steam to exit flue gas and the flue gas recirculation reheat schemes have been evaluated and a report is presently being written. The cyclic-liquid heat exchanger reheat system has been designed and fabricated but not operated.

2. Evaluate the corrosion/erosion resistance of critical scrubber system components as a function of material design and selection. This task tested and evaluated construction material as well as material used in linings. A final report is presently being written.

3. Fabricate, operate, and evaluate new lime/limestone scrubber modes including the horizontal and cocurrent scrubber. The study evaluated gas velocity, liquid circulation rate, solids content of slurry and nozzle pressure drop on SO_2 and particulate removal. Final reports have been written but are presently being edited before publication and distribution.

4. Establish the relationship between the physical and chemical characteristics of scrubber sludge and full-scale scrubber and boiler operation. The project involved the characterization of scrubber sludge from five operating scrubbers. The sludge was analyzed to determine the amount of sulfate present, the degree of thixotropic characteristics exhibited by

the sludge, and the predicted performance of various fixation chemistries possessed by the sludge as a function of the scrubber operation in which it was formed.

RP630, "Evaluation of Improved Process Control Capability for Flue Gas Desulfurization Processes" (Southern California Edison and Stearns-Roger)

The overall objective of this project is to define the best process control methods for lime/limestone scrubbing, from both the standpoint of control design and instrumentation dependability. Specifically, the main objectives are to: (1) define and, if possible, quantity the factors influencing the performance and reliability of chemistry process control for lime/limestone scrubbing; (2) define the state of the art in flue gas desulfurization chemistry process control in terms of these factors; (3) make recommendations on further developments. This project is approximately 75% complete. The final report will be issued about October and will include real-world operating practices and procedures which have been used at full-scale scrubbing facilities. The report will discuss with respect to scrubber control, water balance, pH and SO_2 removal, solids dewatering equipment and factors influencing scale formation.

RP785, "Characterization of Low Sulfur/Alkaline Ash Western Coal for Flue Gas Desulfurization Processes" (Arthur D. Little)

The study has confirmed that simple chemical analysis of a fly ash by titration with an acid does not totally describe how a specific fly ash will react in a scrubber. Investigative efforts indicate a need to develop a highly reliable, accurate and consistent reactivity test to measure the available alkalinity in a fly ash. Presently, coal ash analysis or simple acid titrations are not sufficient to describe how a fly ash will react in a scrubber system. It is imperative that a test or series of tests be developed to accurately reflect how the alkalinity in a specific fly ash will react in a scrubber. Such a test could then be universally applied to develop confident designs for scrubbers planning to use fly ash a scrubbing reagent and could minimize piloting efforts to screen a large number of fly ashes for possible use in a scrubber.

RP786-1-2, "Byproduct/Waste Disposal for Flue Gas Cleaning Processes" (Michael Baker, Jr., and Radian Corporation)

Michael Baker, Jr., and Radian are jointly pursuing the major objective of this project to provide the utility industry a sound data base to permit confident selection of the methods commercially available for disposal of flue gas cleaning waste products. This objective results in the pursuit of four separate tasks.

1. Task 1, recently completed by Radian, reviewed and assessed the existing data base regarding flue gas cleaning wastes. This review and assessment included all data generated during previous and ongoing research and development for sludge and ash disposal. As a result of this task, Radian identified a number of variables affecting the solubility of trace elements, salts and other potentially toxic species in ash and alkaline scrubber sludge that have not been reliably defined in past efforts.

2. As part of Task 2, Radian prepared a test program to quantify the variables that affect the physical and chemical stability of scrubber sludge. During the laboratory test program tests will be performed to measure leachate quality and quantity and the physical strength or load-bearing capacity of sludge and sludge/fly ash blends. It is hoped that these efforts will identify a number of tests that can be utilized to predict scrubber sludge fixation and stabilization properties. Michael Baker, Jr., Inc., is charged with completion of Tasks 3 and 4.

3. In Task 3 Michael Baker will document the state of the art of sludge fixation and determine its practical significance to the utility industry. In completing this task, Michael Baker will establish a data base of major sludge fixation processes commercially developed and potential alternative methods for utility operation. A draft report has been issued concerning this task and is currently being reviewed by EPRI.

4. In Task 4 Michael Baker will, as a result of efforts pursued in the preceding tasks, develop a set of sludge disposal guidelines. The guidelines will be structured in such a manner as to be a useful summary tool

for evaluating alternative approaches and making practical planning decisions. These guidelines will be published toward the end of 1977.

RP786-3, "Sludge Dewatering Equipment" (Envirotech)

The overall objective of this project is to evaluate and define the numerous techniques available for sludge dewatering.

In pursuing the objective of this project, Envirotech is concentrating on evaluating various bench-scale and pilot dewatering devices such as centrifuges, filters and clarifiers. In doing so, Envirotech is attempting to determine what variables in the sludge effect and determine the design of dewatering devices.

RP1031-1, "Study of Alkali Scrubbing Chemistry"

This project, soon to be started, will attempt to quantify the chemical phenomena of lime/limestone slurry scrubbing processes. Much effort has been expended in the evaluation of equipment alternatives such as scrubber type and mist eliminator, but it is clear that the reliability and performance of this equipment is directly related to the chemistry of the scrubbing process. Definition of the chemical phenomena includes interactions between gas, liquid and solids as affected and controlled by slurry and gas composition. Interactions between gas and liquid include gas phase mass transfer and liquid phase mass transfer with chemical reaction and gas/liquid equilibria. Interactions between liquid and solid include dissolution, nucleation, crystal growth and chemical equilibria. All of these interactions are affected by variables such as SO_2 gas concentration, temperature, chloride concentration, Mg concentration, scrubber holdup in contacting vessel, and hold tank residence time.

The main objective of this project is to develop integrated models and correlations of the chemical behavior of lime/limestone scrubbing systems. With these models it is hoped to optimize design and operating variables such as the number of scrubber stages, L/G, limestone utilization and hold tank residence time with respect to the type of scrubbing contactor use (packed bed, turbulent contactor, spray tower, etc.). These models and correlations will make it possible to provide reliable design without large, costly safety factors in equipment size, expected performance and high reagent usage. Using these models to quantitatively evaluate new data will result in potential process improvements involving the use of additives, modified hold tank configurations, modified scrubber vessel design and/or intentional oxidation.

The mathematical models that will be developed under this project will attempt to quantify system performance with respect to:

1. sulfur dioxide removal
2. gypsum supersaturation
3. gypsum subsaturation
4. system oxidation
5. solids quality

RP1033, "Advanced Flue Gas Desulfurization Development and Test Facility" (Tennessee Valley Authority)

This project allows continued operation of TVA's Shawnee Test Facility in Paducah, Kentucky. Three 10 MW test modules have been operated since 1971 at Shawnee under sole EPA sponsorship. EPRI will take over some funding and testing of new equipment at the site beginning in 1978. This project is initially directed at (1) designing and constructing a 10 MW flexible cocurrent/countercurrent spray scrubber module, and (2) modifying the TVA scrubber system to operate in a limestone regenerable double alkali mode. On the EPRI-funded tests TVA would be the primary contractor, with EPRI directing the project in consultation with the established utility Shawnee Advisory Committee. EPRI welcomes EPA participation on this committee. This would insure consideration of the positive technical counsel and experience which the IERL can provide to the utility industry/EPRI funded efforts. We would expect this committee to advise on program needs, review draft test plans and results, as well as insure effective application of the results in a technical transfer program. In turn, we would be pleased to cooperate in a similar advisory capacity on the continuing EPA efforts at Shawnee. In each case, however, the organizations funding a particular project should have technical and financial decision authority over that project.

Regenerable Flue Gas Desulfurization

In line with the specific objectives of the SO_x control subprogram, EPRI's work in regenerable flue gas desulfurization has been established to complement the lime/limestone work. It provides options for improved SO_2 removal, reduced water pollution and solid waste impact, reduced raw material use, and lower-cost systems at specific sites. The EPRI development approach is progressing from (1) process comparison and technical evaluation to (2) detailed site-specific design and then to (3) construction and testing of appropriately-sized prototype facilities. In addition, EPRI is evaluating full-scale demonstration plants in conjunction with the host utilities.

RP535, "Evaluation of Advanced Regenerable Flue Gas Desulfurization Processes (Radian Corporation)

This project reviewed emerging processes and evaluates 12 systems, including three that have been tested on a large scale (Wellman-Lord, magnesia, and Cat-Ox) and lime/limestone as a baseline. The report evaluates sludge regeneration from lime/limestone scrubbing sulfur versus sulfuric acid production, and subloop gasifier problems. It has chemical, mass, and energy balances for the processes and puts the systems on a common design basis. Operating costs were calculated, but detailed capital costs were not included because the processes differ widely in their state of development and the comparative equipment costs would not be realistic without more detailed estimates.

Some general results of the evaluation are:

- Calcium sulfate/sulfite sludge regeneration is not currently practical due to cost, energy use, and chemical complexity.
- The magnesia and Wellman-Lord systems are further developed and are not inherently inferior to other less-developed regenerable processes.

- The less-developed processes have specific advantages in various subsystems of the process, but these processes require more development at pilot or prototype scale before being confidently installed commercially.

Each FGD system under evaluation has unique design features that present potential advantages and problems in acceptance of the process by the electric utility industry. A list of FGD processes, their development status, advantages, and concerns is shown in Table 2.

Detailed information, including flow sheets, chemical energy balances, information on sulfur versus sulfuric acid production, gasification subsystems, and additional process information, is available in the two-volume final report, EPRI FP 272, Vol. 1 and 2 (currently being reprinted).

RP536, "Evaluation of Three Prototype Flue Gas Desulfurization Processes" (Southern Company Services)

Three 20 MW advanced prototype FGD systems were tested at the Scholz Plant of Gulf Power. EPRI supported part of this test work and will issue the final report shortly. The three systems are Chiyoda 101, Combustion Equipment Associates/Arthur D. Little (CEA/ADL) Dual Alkali, and Bergbau Forschung/Resox processes. Chiyoda 101 scrubs with dilute sulfuric and externally reacts this with limestone to produce commercial gypsum. While the system was plagued with several mechanical problems (centrifuges, FRP oxidizer and piping) and a fire, the overall process performance was excellent during all phases of operation. Process limitations of large equipment size, high L/G, high auxiliary power, and stainless steel construction may be overcome in the Chiyoda 121 process, which is now being considered for follow-on work at Scholz.

The Combustion Equipment Associates/Arthur D. Little, Inc., process is a sodium-based concentrated mode dual alkali system with lime regeneration and sulfite/sulfate sludge production. The most important limits are its requirement of lime and makeup sodium and the sludge character with sodium and sulfite leaching potential. Although there were several mechanical problems with the system, particularly the filter system and corrosion-resistant liner

TABLE 2. EVALUATION SUMMMARY OF REGENERABLE FGD PROCESSES
RESULTS OF RP535 - RADIAN CORPORATION

PROCESS	COMPANY	DEMONSTRATION SIZE (MW)	MAIN ADVANTAGES	REMAINING CONCERNS
Carbon Adsorption				
Moving Bed[1]	Bergbau-Forschung Foster-Wheeler Energy Corp.	42	Dry; uses coal as direct reductant; prototyped	Much solids handling; hot spots in adsorber; proving RESOX
Fluidized Bed[2]	Westvaco	0.2	Dry; relatively simple	Fluid-bed operation; sulfate emission; requires H_2; absorbent attrition
Copper Oxide[1]	Shell-UOP	40	Dry; very good for NO_x; prototyped	High capital cost; highest energy consumption; requires H_2; mechanically complex
Sodium Salt Scrubbing				
Thermal Regeneration (Wellman-Lord)	Davy Powergas/Allied Chemical	220	Good SO_x removal; well-developed, simple operation of absorber	Wet; sulfate waste produced, gaseous reductant required
High-Temperature Reduction[2] (ACP)	Atomics International Div., Rockwell Intern'l Corp.	1-2	Semi-dry; uses coal as direct reductant	Delicate absorber control; high temperature in reducer; particulate control
Electrolytic Regeneration[2]	Ionics	0.75	May provide backfit sodium regeneration	Wet; gaseous reductant required; high power consumption; weak sulfuric acid as byproduct
Citrate as Buffer (or phosphate)[2]	Pfizer, Inc.; Peabody Engineering Corp.; Chemico Division, Envirotech Corp.	1	One-step sulfur production; no concentrated sulfur dioxide stream as a gas	Difficult product separation; hydrogen sulfide required as reductant; wet
Ammonia Scrubbing				
Thermal Regeneration; Sulfur Dioxide Reduction[2]	Cottrell Environmental Research Cornell, Inc. Catalytic, Inc./I.F.P.	30	Excellent sulfur dioxide removal; prototyped; proven sulfur dioxide reduction	Fume potential; syngas required as reductant; wet
Bisulfate Regeneration[1]	Tennessee Valley Authority	1.2	Low liquid-to-gas ratio; good absorption	Fume potential; bisulfate very corrosive; wet reducer operation

TABLE 2 (continued)

PROCESS	COMPANY	DEMONSTRATION SIZE (MW)	MAIN ADVANTAGES	REMAINING CONCERNS
Magnesia Scubbing[3] (Mag-Ox)	United Engineers and Constructors, Inc.; Chemico Division, Envirotech Corp.; others	160	Relatively simple; prototyped	High-temperature kiln operation; wet; loss of reactant; reliability problems
Catalytic Oxidation[3] (Cat-Ox)	Monsanto Company	110	Simple; low labor and material cost; dry	High cost, sulfate emission; hot electrostatic precipitator required; mechanical operation is complex
Lime/Limestone Scrubbing				
Calcium Sulfite-Sulfate Reduction[3]	Various	(Absorber) 800+	Known scrubbing; regeneration loop can retrofit	Wet; high-temperature, high-cost regeneration

[1] Either sulfur or sulfuric acid can be produced as byproduct
[2] Only sulfur as byproduct
[3] Only sulfuric acid as byproduct

systems used to protect the carbon steel vessels, overall system performance was excellent during all phases of operation. Follow-on work at Scholz involving limestone rather than lime for regeneration is also under consideration.

The Bergbau Forschung/Foster Wheeler Resox process at Scholz adsorbs SO_2 and promotes oxidation to sulfuric acid on char pellets, uses hot sand in a fluid bed to reconvert to a concentrated SO_2 stream, and then reacts the SO_2 with coal in the Resox reactor to form elemental sulfur. The process had debilitating mechanical problems and never had a sufficiently long sustained run of integrated operation to confirm performance reliability. The mechanical design is complex and differs from the German configuration. This U.S. design would require major modification in order to achieve operational success.

RP784-1, "Detailed Design and Evaluation of Advanced Flue Gas Desulfurization Processes" (Stone and Webster Engineering Corporation)

The objective of this project is to develop and evaluate integrated system designs of advanced regenerable flue gas desulfurization processes. Steps in the project were: select engineering firms (Stone and Webster); issue process proposal request (to 14 vendors); evaluate 10 proposals, including technical and economic assessment. The detailed design was expected to be for several competing processes. The economic comparisons and technical assessment gave an unexpected recommendation, which was to go ahead with only one detailed design -- Foster Wheeler Resox. Additionally, the Peabody and Spring Chemical proposals showed process potential, but lacked technical supporting information and development work. A 500 MW capital and operating cost comparison was developed to choose the most attractive processes for further development. Follow-on efforts will refine unsubstantiated vendor costs and will add capital and operating costs for lime/limestone, AI Aqueous Carbonate Processes, Wellman-Lord and magnesia scrubbing to the cost estimates. A final report on the evaluation for economic comparison will be published after the follow-on efforts are complete.

RP784-2, "Regenerable Flue Gas Desulfurization - Resox" (Foster Wheeler Energy Corporation)

The recommendaton in RP784-1 led to EPRI authorizing the design, construction, operation and testing of a 42 MW size Resox unit at the Lünen, West Germany, plant of Steag A.G. The Resox unit reacts concentrated SO_2 with coal to form sulfur. This plant has an operating Bergbau-Forschung unit and gas blending facility that can simulate alternate process "front end" feeds of rich SO_2. The German government agency Umsweltbundesamt (UBA) is program sharing in operating costs of the Bergbau-Forschung (BF) front end. We will receive detailed information on this operating BF unit. Advantages of this project include: (1) completion of one year's testing by mid-1979, (2) operation of the Resox on an operating coal-fired boiler/FGD process front end. Institute research objectives at Lünen are to: (1) prove sustained operation at 42 MW scale, (2) confirm Resox design, (3) provide mixed gas compositions similar to Wellman-Lord, MgO and absorption/steam stripping rich gas to demonstrate flexibility, (4) demonstrate load-following capability of Resox.

RP981, "Aqueous Carbonate Flue Gas Desulfurization Process Development" (Atomics International)

This project supports the EPA-ESEERCO Aqueous Carbonate Process (ACP) demonstration plant and tests key design areas. This process uses a spray dryer to contact sodium carbonate with flue gas. Spent sulfite is then reduced with coal by going through several severe processing steps, including a molten salt bath, melt dissolution and ash precipitation steps. EPA and ESEERCO plan an ACP demonstration at Huntley Station of Niagara-Mohawk. This EPRI support project with Niagara-Mohawk will test melt tapping and quench, solution chemistry and ash dissolution/reprecipitation, and component corrosion. The tests will be done in AI's 5 MW facility at Canoga Park, California. Key questions to be addressed are the system ability to handle coal, the tapping, quenching and dissolution of a salt/ash melt, and the ability to operate liquid chemical loops that contain silica and iron compounds and precipitate ash out of solution. This project has been authorized for some time but contract negotiations have delayed start of work until late 1977. Work should be complete in Spring 1978.

RP1032, "Testing and Evaluation of Magnesium Oxide Regenerable FGD Demonstration Unit" (Philadelphia Electric)

The only operating magnesia (MgO) scrubber in the U.S. will be evaluated in six modes with special testing and instrumentation added to characterize the scrubber. Solids will be analyzed for crystal form, waters of hydration and chemical composition. Liquid streams will be analyzed for major and trace elements. Gas streams will be tested for sulfates, chlorides and normally monitored gases; chemical and mass balances will be performed. EPRI's participation will add special testing and report the results from this leading regenerable FGD process to the utility industry. This contract is near signature. A final report should be available in mid-1978.

RP491-1, "Evaluation of Dry Alkalis for Removing Sulfur Dioxide from Boiler Flue Gases" (Bechtel Corporation)

Bechtel reviewed the status of dry alkali injection for SO_2 removal for EPRI and this evaluation is published in report, EPRI FP 207 October 1976. Dry injection tests, particularly of nahcolite, are reported as well as related information such as chemistry, mining, tramsportation, waste disposal and economics. Discussion of the Wheelabrator-Frye and Superior tests are a large part of the report.

RP982-7, "Laboratory Evaluation of Dry Alkalis for Dry SO_2 Removal" (KVB)

This project will obtain well characterized data that will define the dependence of the dry alkali SO_2 removal process on various system parameters. These parameters will include several sorbents, temperature-time history, amount of SO_2 removed in suspension versus removal on baghouse filter cake, effect of stoichiometric ratio and any byproduct emissions. The results help define the larger-scale tests now contemplated for the EPRI Arapahoe Particulate Test Facility in Denver, Colorado. This contract is in negotiation.

CONCLUSION

EPRI has increased its efforts in evaluating, reporting, testing and developing new technologies for flue gas desulfurization. Reports that cover survey and evaluation work have been published. Results from several pilot and demonstration test programs are now in draft form and will be published shortly. We are moving into demonstration projects at Shawnee Station of TVA in lime/limestone and dual alkali, at Lünen, West Germany, in regenerable FGD subsystems, and in dry SO_2 removal at the EPRI Particulate Test Facility. New processes do not need to be developed for novelty's sake, but utilities do need economic and reliable choices that will fit their site, coal type, waste product disposal requirements, unit size and load pattern in the 1980-1990 time frame. We feel EPRI can play an important part in developing and clarifying these process choices.

ECONOMIC EVALUATION TECHNIQUES, RESULTS, AND COMPUTER MODELING FOR FLUE GAS DESULFURIZATION

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ABSTRACT

The Emission Control Development Projects group of the Tennessee Valley Authority (TVA), Office of Agricultural and Chemical Development, is involved in evaluating comparative economics of alternate SO_2 control technologies. Given within are the techniques and results of recent U.S. Environmental Protection Agency (EPA) sponsored economic evaluations for a U.S. Bureau of Mines-type citrate process and a generic double-alkali process utilizing updated design and economic premises. For comparison, an updated evaluation for limestone scrubbing utilizing similar design and economic premises is given.

In conjunction with the EPA-sponsored Shawnee test program, Bechtel Corporation and TVA have jointly developed a computer program capable of projecting material balances and relative economics for lime and limestone scrubbing utilizing a Turbulent Contact Absorber. The current program may also be utilized to project the effect of variations in lifetime operating profile and alternative sludge disposal methods on costs for either process. Although the program is not intended to compute the economics of an individual system to a high degree of accuracy, it allows prospective users to quickly project comparative design and costs for various limestone-lime case variations on a common design and cost basis. The technical and economic bases for the current program are presented. In addition, future process design alternatives to be incorporated into the program and limitations of the program validity are discussed.

ECONOMIC EVALUATION TECHNIQUES, RESULTS, AND COMPUTER MODELING FOR FLUE GAS DESULFURIZATION

INTRODUCTION

Under contract to the U.S. Environmental Protection Agency (EPA), the Tennessee Valley Authority (TVA) Emission Control Development Projects group (ECDP) is involved in several economic evaluations of alternative methods for sulfur (S)-sulfur oxides (SO_x) control and for disposal of byproduct-related wastes. One major evaluation program is subdivided into four tasks including (1) the selection, definition, and economic evaluation of up to seven advanced sulfur dioxide (SO_2) removal processes, (2) the technical and economic evaluation of various methods for disposal of lime-limestone scrubbing wastes, (3) the study and evaluation of several front-end, fuel-cleaning technologies, and (4) the review of process evaluation reports prepared by others. A second major evaluation program includes the development of a computer model for projecting comparative investment and revenue requirements for lime and limestone scrubbing systems based on the results of the EPA-TVA-Bechtel Shawnee scrubbing test facility.

In the first program mentioned above, two advanced flue gas desulfurization (FGD) processes have recently been evaluated including a generalized or generic double-alkali system and a citrate process. Recently updated technical and economic premises and the results of these evaluations are described in this paper. Because the majority of FGD installations in the U.S. currently utilize limestone-lime scrubbing, an updated limestone slurry process is also included for comparison with the two advanced systems mentioned. Results of a companion task evaluating sludge disposal options are to be presented in another paper during the symposium.

For the second evaluation program mentioned, details of the Shawnee limestone-lime computer model are discussed including the present and future program scope and the process definition for the current options.

FCD EVALUATION TECHNIQUES

TVA has been involved in the preparation of technical and economic evaluations of alternate FGD processes for EPA, and others, since 1967. Many of these evaluations have been published; perhaps the best known is Detailed Cost Estimates for Advanced Effluent Desulfurization Processes (EPA-600/2-75-006; PB 242 541/1WP, January 1975). Since this publication, design and economic premises for future evaluations have been modified through discussion with EPA and others to reflect prevailing fuel characteristics, current design practice, and economic conditions. Given below are the updated premises on which the advanced regenerable process evaluations are based.

Design

Base Case. The base case for conceptual design and detailed cost estimating is a 500-MW new power unit with a heat rate of 9000 Btu/kWh, burning 3.5% S coal (dry basis). The coal composition has been adjusted from the initial study reflecting a reduced heating value (HHV) of 10,500 Btu/lb (as fired) and a higher average ash content of 16%. The as-fired coal composition and flow rate for the base case design is shown below.

Coal composition (3.5% S, dry basis)	Wt %, as fired	Lb/hr[a]
C	57.56	246,800
H	4.14	17,700
N	1.29	5,500
O	7.00	30,000
S	3.12	13,400
Cl	0.15	600
Ash	16.00	68,600
H_2O	10.74	46,000
Total	100.00	428,600

a. 500-MW new unit with a heat rate of 9000 Btu/kWh.

Operating life. The projected operating life of a new coal-fueled power unit is assumed to be 30 yr representing a total generation of 127,500 hr of generating capacity over the life of the plant. The projected load factor for the first year of operation is assumed to be 7000 hr of generating capacity.

Flue gas composition. Flue gas composition is based on the combustion of pulverized coal assuming a total air rate to the air preheater equivalent to 133% of the stoichiometric requirement. This includes 20% excess air to the boiler and 13% air inleakage at the air preheater. A horizontal, frontal-fired, coal-burning unit is assumed. For this unit, it is assumed that 80% of the ash present in the coal is emitted as flyash and 95% of the S in the coal is emitted as SO_x. One percent of the S emitted as SO_x is assumed to be sulfur trioxide (SO_3) and the remainder SO_2. The S content of coal and the corresponding SO_2 emission rates are based on average values and do not take into consideration variations which may be encountered in actual coal deliveries. Actual S contents are reported to be as much as 22% greater than average values. The basis for calculating the flue gas rates and compositions are tabulated below.

Flue Gas Combustion Calculation Basis

Power unit size, MW	500
Power unit heat rate, Btu/kWh	9,000
S content of coal, wt % (dry basis)	3.5
Heating value of coal, Btu/lb (as fired, HHV)	10,500
Coal to power unit, lb/hr	428,600

Flue Gas Composition and Properties

Component	Vol, %	Lb/hr
N_2	73.76	3,450,000
O_2	4.83	258,200
CO_2	12.31	904,200
SO_2	0.24	25,130
SO_3	0.0024	317
NO_x	0.06	3,009
HCl	0.01	661
H_2O	8.79	264,500
		4,906,000 lb/hr (approx)
		1,543,000 aft^3/min at 300°F (approx)

Flyash loading, gr/sft^3 (60°F) dry basis 6.65

Flyash loading, gr/sft^3 (60°F) wet basis 6.06

Particulate removal. Electrostatic precipitators (ESP) designed to remove 99.5% of the particulates are used for each process; however, costs for the ESP are not included in the estimates. Each scrubbing system is designed with a humidification chamber upstream of the scrubbers. It is assumed that 5% of the SO_2, 50% of the SO_3, and 100% of the hydrogen chloride (HCl) in the flue gas will be removed from the gas during humidification. These compounds are discarded along with the byproduct solids in throwaway processes. In processes that produce salable products, these compounds are neutralized and discarded separately.

Degree of SO_2 removal. For the processes presented here, SO_2 removal is based on meeting the current SO_2 emission regulation of 1.2 lb SO_2 allowable emission/MBtu (M = one million) heat input. For the base case coal composition, an SO_2 removal efficiency of 78.8% is required. A 22% increase in the S content of delivered coal would require that the SO_2 removal efficiency be increased to 82.6%.

Redundancy. Because process design is based on assumed demonstration level technology, no special redundancy is provided and only pumps are spared. The base case does not include a bypass around either the ESP or the FGD unit.

Reheat. Indirect steam reheat is used for all cases. Entrainment is estimated as 0.1% of the wet gasflow rate at the scrubber outlet; however, in actual practice this value may range from 0.1 to 0.5%.

Waste disposal. For the double-alkali and limestone processes an onsite disposal pond lined with impervious clay is used. The pond is assumed to be located 1 mi from the scrubbing site.

Investment

Plant location. A midwestern location has been chosen for estimating capital investment because of the concentration of electrical utilities and the proximity of coal fields in that area.

Project schedule. Projects are assumed to begin in mid-1977 and end in mid-1980, with an average cost basis for scaling of mid-1979. Direct investments are prepared using the average annual *Chemical Engineering* cost indexes and the TVA projections shown below.

Cost Indexes and Projections

Year	1973	1974	1975	1976[a]	1977[a]	1978[a]	1979[a]	1980[a]	1981[a]
Plant	144.1	165.4	182.4	197.9	214.7	232.9	251.5	271.6	293.3
Material[b]	141.9	171.2	194.7	210.3	227.1	245.3	264.9	286.1	309.0
Labor[c]	157.9	163.3	168.6	183.8	200.3	218.3	237.9	259.3	282.6

a. Projections.
b. Same as index in *Chemical Engineering* for "equipment, machinery, supports."
c. Same as index in *Chemical Engineering* for "construction labor."

Direct investment basis. The direct investment estimate begins with the inlet plenum downstream from the ESP and includes all of the equipment between the inlet plenum and the stack plenum following the reheater. Particulate removal and disposal equipment and costs are not included in the process investment estimates.

Services, utilities, and miscellaneous costs are estimated as 6% of the process areas subtotal. This covers such items as maintenance shops, stores, and communication, railroad, and fire and service water facilities.

Indirect investment basis. In addition to the direct costs which include materials and labor for equipment and installation, services and utilities, and pond construction, the indirect costs for the project (which include in-house engineering design and supervision, architect and engineering contractor expenses, contractor fees, and construction expenses) are included in the investment estimates. Construction facilities, which include costs for mobile equipment, temporary lighting, construction roads, raw water supply, safety and sanitary facilities, and other similar expenses incurred during construction, are considered a part of construction expenses. Consultant fees are not included. The engineering design and supervision, and the contingency factors are based on demonstration-level technology and experience. Indirect investment costs are estimated from the number of drawings required, man-hours of supervision and construction, etc.

Allowances. Allowances are included for startup and modification, interest during construction, and working capital. Startup and modification allowances are estimated as 10% of total fixed investment for the recovery process and 10% of the total fixed investment minus pond construction for the throwaway processes. Interest during construction is estimated as 12% of the subtotal fixed investment for each process. This factor is equivalent to the simple interest which would be accumulated at a 10%/yr rate assuming an incremental capital structure of 60% debt, 40% equity, and a 3-yr project expenditure schedule as indicated below.

Project Expenditure Schedule

	Year			
	1	2	3	Total
Fraction of total expenditure as borrowed funds	0.15	0.30	0.15	0.60
Simple interest at 10%/yr as % of total expenditure				
Year 1 debt	1.5	1.5	1.5	4.5
Year 2 debt	-	3.0	3.0	6.0
Year 3 debt	-	-	1.5	1.5
Accumulated interest as % of total expenditure	1.5	4.5	6.0	12.0

Working capital. Working capital consists of the total amount of money invested in raw materials and supplies carried in stock, finished products in stock, and semifinished products in the process of being manufactured; accounts receivable; cash kept on hand for payment of operating expenses such as salaries, wages, and raw material purchases; accounts payable; and taxes payable. For these premises, working capital is defined as the equivalent cost of 3 wk of raw material costs, 7 wk of direct costs, and 7 wk of overhead costs.

Revenue Requirements

Direct costs. Annual revenue requirement display tables are based on 7000 hr of operation/yr. Process operation schedules are assumed to be the same as the power plant operating profiles. Raw material, labor, and utility costs are projected to 1980. Maintenance costs are estimated on the basis of direct investment and are varied for each process as a function of unit size.

Indirect costs. Following power industry practice, regulated company economics are used for establishing capital charges. The projected breakdown for annual capital charges is shown below.

Annual Capital Charges for Power Industry Financing

	Percentage of total depreciable capital investment		
	Years remaining life		
	30	25	20
Depreciation-straight line (based on years remaining life of power unit)	3.3	4.0	5.0
Interim replacements (equipment having less than 30-yr life)	0.7	0.4	-
Insurance	0.5	0.5	0.5
Property taxes	1.5	1.5	1.5
Total rate applied to original investment	6.0	6.4	7.0

	Percentage of unrecovered capital investment[a]
Cost of capital (capital structure assumed to be 60% debt and 40% equity)	
Bonds at 10% interest	6.0
Equity[b] at 14% return to stockholder	5.6
Income taxes (Federal and state)[c]	5.6
Total rate applied to depreciation base	17.2[d]

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a. Original investment yet to be recovered or "written off."

b. Contains retained earnings and dividends.

c. Since income taxes are approximately 50% of gross return, the amount of taxes is the same as the return on equity.

d. Applied on an average basis, the total annual percentage of original fixed investment for new (30-yr) plants would be 6.0% + 1/2 (17.2%) = 14.6%.

In estimating the regulated capital charges associated with stack-gas scrubbing, the conventional method of considering the overall life of the power plant is used. The Federal Power Commission (FPC) recognizes the conclusion of the National Power Survey that a 30-yr service life is reasonable for steam-electric plants. Because some items have life spans of less than 30 yr, however, FPC has designated interim replacements as an allowance factor to be used in estimating annual revenue requirements to provide for the replacement of such items. Use of this allowance following FPC recommended practice provides for financing the cost of replacing such short-lived units. An average allowance of about 0.35% of the total investment is normally provided. However, to provide for the unknown life span of SO_2 control facilities, a somewhat larger allowance factor which is varied with projected plant life is used for new units. An insurance allowance of 0.5% of total depreciable capital investment is also included in the capital charges based on FPC practice. Property taxes are estimated as 1.5% of the total depreciable capital investment.

Debt to equity ratio is another component of capital charges for which variations of ratios may be expected. FPC data indicate that the long-term debt for privately owned electric utilities varied only slightly from 51.5 to 54.8% of total capitalization during the period 1965-73. However, recent economic upheavals have changed the incremental debt to equity ratio as utilities are forced to depend more and more on bonds and bank loans for project funding. For this study a debt to equity ratio of 60:40 is assumed. Cost of capital and income tax charges are applied to the unrecovered portion of capital investment.

Since most regulatory commissions base the annual permissible return on investment on the remaining depreciation base (that portion of the original investment yet to be recovered or written off), a portion of the annual capital charge included in the lifetime operating costs declines uniformly over the life of the power plant.

Overheads. Plant, administrative, and marketing overheads are costs which vary from company to company. With consideration of the various methods used in industry and illustrated in a variety of cost estimating sources, the following method for estimating overheads is used.

Plant overhead is estimated as 50% of the subtotal conversion costs less utilities, which includes the projected costs for labor, maintenance, and analyses. Administrative overhead is estimated as 10% of operating labor and supervision. Marketing the product is considered in the estimation of overheads and is defined as 10% of sales revenue.

Byproduct sales. In the evaluation of annual and lifetime economics, credit from sale of byproducts is deducted from the yearly projections of operating cost to obtain the net effect of the pollution process on the cost of power.

FLUE GAS DESULFURIZATION EVALUATION RESULTS

Descriptions of the three processes evaluated are presented below along with the projected investment and revenue requirement results.

Process Descriptions

Limestone slurry process. The limestone slurry process for desulfurization of flue gas (Fig. 1) assumes flyash removal by ESP. A common plenum is placed downstream from the ESP and the power plant induced-draft (ID) fans to distribute the gas to the absorbers. Forced-draft booster fans, relative to the absorbers, are provided on the downstream side of the plenum to overcome the pressure drop in the FGD system.

Makeup limestone slurry from the feed preparation area is combined with scrubber effluent slurry and recycle pond water to control the concentration of the recirculating slurry at approximately 15% solids. The flue gas is cooled in a humidification chamber and fed to the mobile-bed absorbers. The limestone slurry circulates through the absorbers where it reacts with the SO_2 in the cooled flue gas. The absorbers are equipped with chevron-type entrainment separators with

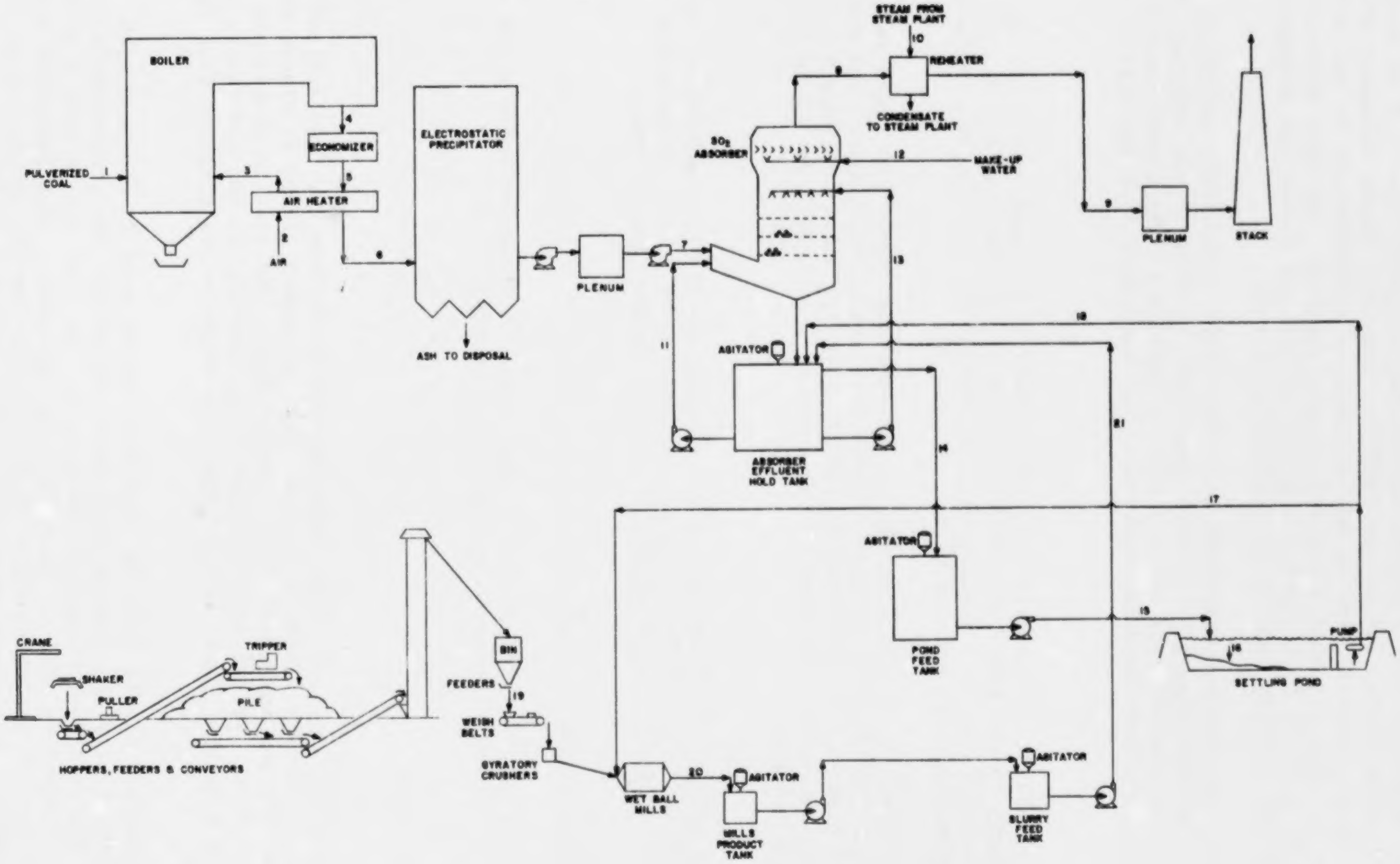

Figure 1. Limestone slurry process.

provisions for upstream and downstream wash with fresh makeup water to control entrainment carryover in the gas stream. A bleed stream from the effluent hold tank is fed to the pond feed tank and the spent slurry is pumped to the onsite pond where the solids in the slurry settle to form a sludge containing approximately 40% solids. Pond supernate is recycled to the wet ball mills and the absorber effluent hold tank to maintain closed-loop operation. Scrubber outlet gas is reheated to 175°F by indirect steam heat before entering the stack.

Generic double-alkali process. The double-alkali process included in this study (Fig. 2) has been generalized from the several processes currently offered in the U.S. For this process, an ESP is used for removal of flyash and a common plenum and booster fans are included downstream from the ESP and the power plant ID fans for distribution of the gas. Flue gas is cooled and saturated in a humidification chamber by a recycle stream of scrubber effluent. In the absorber tower SO_2 is removed using a regenerated sodium solution and additional recycle scrubber effluent. The outlet gas from the scrubber passes through a chevron-type mist eliminator with provisions for upstream wash with fresh makeup water. The cleaned flue gas is reheated to 175°F by indirect steam heat before entering the stack.

Pebble lime is slaked and then reacted with a bleed stream of absorber effluent in agitated tanks. The reaction product, predominately calcium sulfite ($CaSO_3 \cdot 1/2H_2O$), flows to a thickener where the slurry is concentrated. This stream is further dewatered using drum filters to produce a cake containing 60-80% solids. The filter is designed with two wash sections to minimize sodium loss. Filter cake is reslurried with supernate from the pond and fresh makeup water for pumping to disposal. The solids settle to a concentration of approximately 40% in the pond. The required makeup sodium carbonate (Na_2CO_3) is added to the regenerated scrubber liquor at the thickener overflow storage tank.

Citrate process. The citrate process design developed for this study (Fig. 3) is adapted from a U.S. Bureau of Mines (BOM) process. The scheme evaluated assumes flyash removal by ESP. A common plenum and booster fans downstream from the ESP and the power plant ID fans are included in the design.

The flue gas is first cooled in a humidification chamber. SO_2 is then removed from the saturated flue gas by countercurrent scrubbing in a packed tower using a regenerated solution of sodium citrate as a buffer. A purge stream to control chlorides is pumped from the bottom of the absorber through a stripper to a neutralization tank where it is reacted with lime before being pumped to the ash disposal pond. Cleaned flue gas is passed through a chevron-type entrainment separator with provisions for upstream wash with fresh makeup water and reheated to 175°F by indirect steam heat before entering the stack.

Elemental S is precipitated from the SO_2-laden sorbent by countercurrent contact with hydrogen sulfide (H_2S) gas containing approximately 80-97% H_2S. The S is separated by flotation, then melted and settled from the slurry liquor in a decanter operating at a pressure of about 35 psi.

About 2% of the absorbed SO_2 is oxidized in the decanter to sodium sulfate (Na_2SO_4) which is removed from the recirculated sorbent by crystallization as byproduct Glauber's salt ($Na_2SO_4 \cdot 10H_2O$). This sodium is replaced by a mixture of sodium hydroxide (NaOH) or Na_2CO_3 and citric acid. The system guards against H_2S

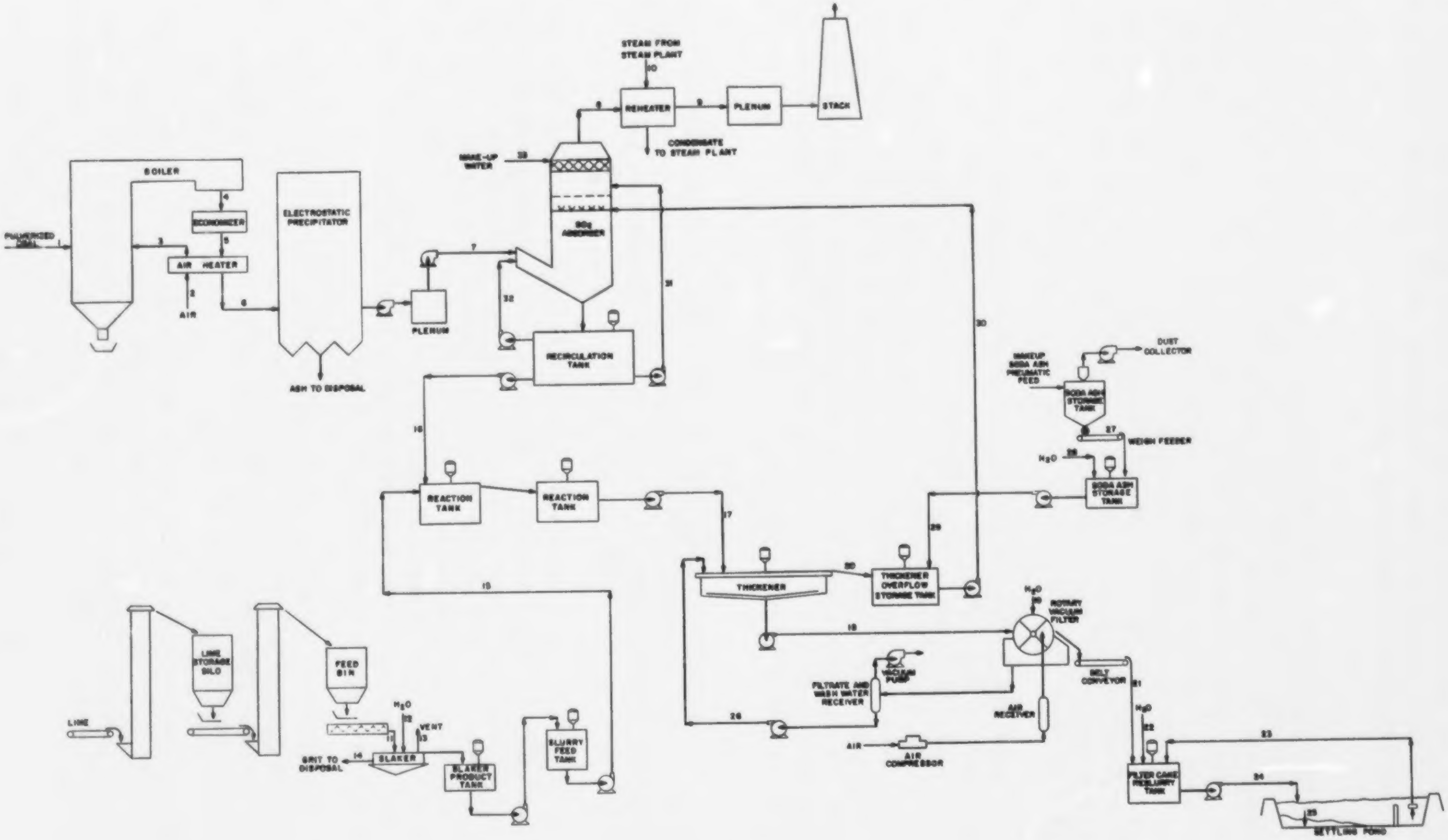

Figure 2. Generic double-alkali process.

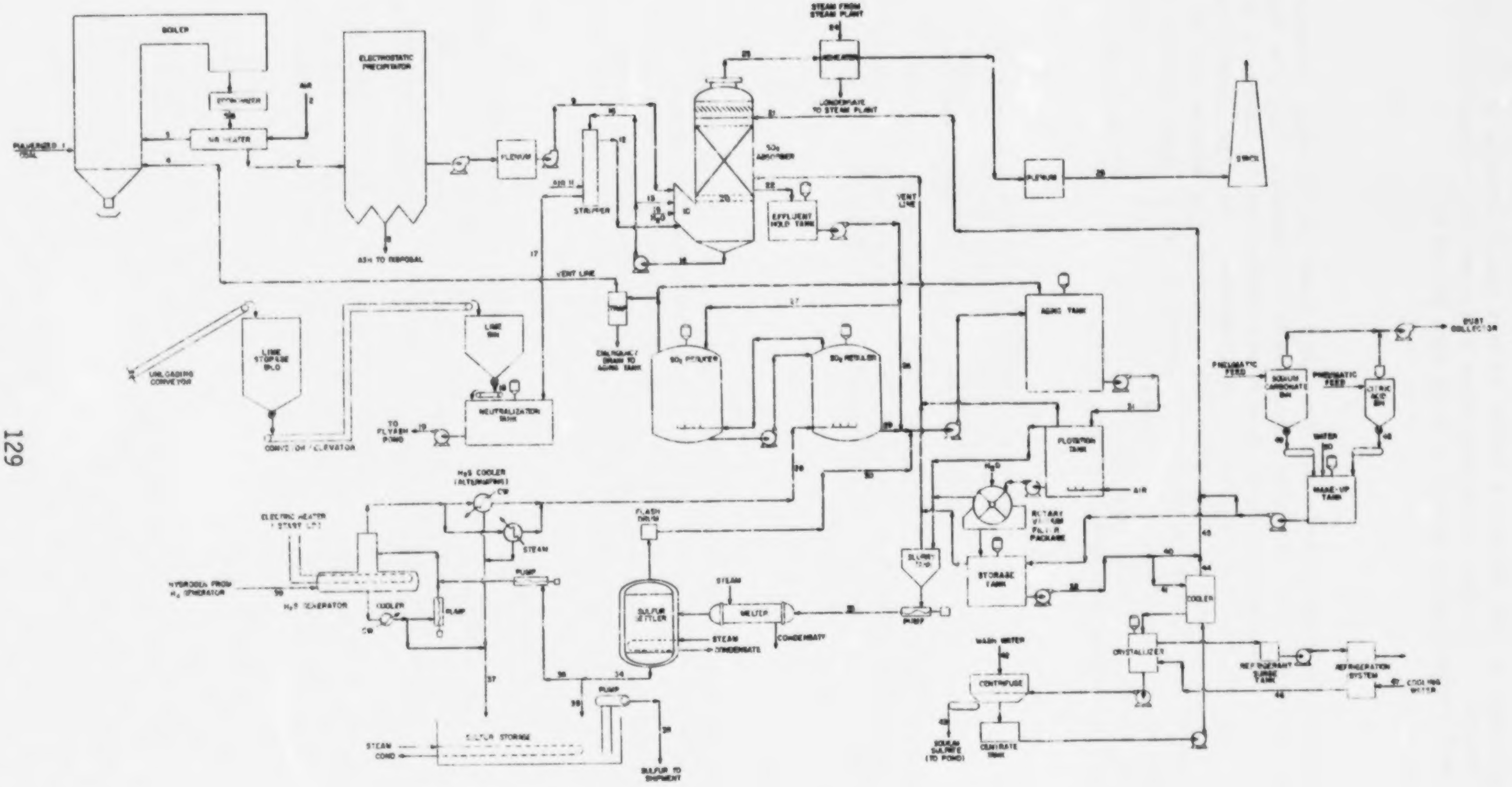

Figure 3. Citrate process.

escape by returning unutilized H_2S to the boiler for incineration and by neutralizing dissolved H_2S downstream of the reducing tanks with a small stream of SO_2-laden liquor from the absorber. The H_2S generator uses hydrogen (H_2) from natural gas or coal gasification and recirculated molten S from the decanter.

Investment

Using the premises outlined previously for base case conditions, investment summaries have been estimated for the three processes discussed. For the limestone slurry process, installation expenses are determined individually based on detailed layout drawings and projected erection labor requirements. For the generalized citrate and double-alkali processes, installation expenses are determined as a percentage of equipment cost based on projected erection labor requirements from the literature for comparable type equipment. Summaries of the projected investment requirements for the three processes are shown in Tables 1-3. Comparisons of the direct and total capital investments are given below.

	Investment, 1979 $		
	Total direct investment	Total capital investment	$/kW
Limestone slurry	26,000,000	48,728,000	97
Generic double alkali	26,750,000	50,551,000	101
Citrate	38,838,000	74,889,000	150

Revenue Requirements

First-year revenue requirements are estimated for each of the three processes. The results, based on 7000 hr of operation, are displayed in Tables 4-6. Given below is a comparison of the total annual revenue requirements and equivalent unit revenue requirements for the three processes expressed in mills/kWh, $/ton coal burned, $/MBtu heat input, $/short ton of S removed, and $/short ton of S produced (citrate process).

		Equivalent unit revenue requirements, 1980$				
FGD system	Total annual revenue requirements, $	Mills/ kWh	$/ton coal burned	$/MBtu heat input	$/short ton S removed	$/short ton S sold
Limestone slurry	14,100,900	4.03	9.40	0.45	408.01	-
Generic double alkali	14,675,000	4.19	9.78	0.47	424.62	-
Citrate	23,264,200	6.65	15.51	0.74	673.15	687.48

Table 1. LIMESTONE SLURRY PROCESS

SUMMARY OF ESTIMATED FIXED INVESTMENT[a]

(500-MW new coal-fired power unit,
3.5% S in coal; 1.2 lb SO_2/MBtu
heat input allowable emission; onsite solids disposal)

	Investment, $	Percent of total direct investment
Direct Investment		
Materials handling (hoppers, feeders, conveyors, elevators, bins, shaker, puller)	1,759,000	6.8
Feed preparation (feeders, crushers, ball mills, hoist, tanks, agitators, and pumps)	1,740,000	6.7
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack)	4,318,000	16.6
Absorption (4 TCA scrubbers including humidification chambers and mist eliminators, effluent hold tanks, agitators, and pumps)	8,918,000	34.3
Stack gas reheat (4 indirect steam reheaters)	1,282,000	4.9
Solids disposal (onsite disposal facilities including feed tank, agitator, slurry disposal pumps, and pond water return pumps)	1,658,000	6.4
Subtotal	19,675,000	75.7
Services, utilities, and miscellaneous	1,180,000	4.5
Total process areas excluding pond construction	20,855,000	80.2
Pond construction	5,145,000	19.8
Total direct investment	26,000,000	100.0
Indirect Investment		
Engineering design and supervision	1,207,000	4.6
Architect and engineering contractor	268,000	1.0
Construction expense	3,617,000	13.9
Contractor's fees	1,142,000	4.4
Total indirect investment	6,234,000	23.9
Contingency	6,447,000	24.8
Total fixed investment	38,681,000	148.7
Other Capital Charges		
Allowance for startup and modifications	3,354,000	12.9
Interest during construction	4,652,000	17.9
Total depreciable investment	46,677,000	179.5
Land	1,030,000	4.0
Working capital	1,021,000	3.9
Total capital investment	48,728,000	187.4

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[a] Basis:
Midwest plant location represents project beginning mid-1977, ending mid-1980.
Average cost basis for scaling, mid-1979.
Stack gas reheat to 175°F by indirect steam reheat.
Minimum in-process storage; only pumps are spared.
Disposal pond located 1 mi from power plant.
Investment requirements for flyash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 2. GENERIC DOUBLE-ALKALI PROCESS

SUMMARY OF ESTIMATED FIXED INVESTMENT[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; onsite solids disposal)

	Investment, $	Percent of total direct investment
Direct Investment		
Materials handling (conveyors, elevators, bins, and feeders)	1,667,000	6.2
Feed preparation (feeders, slakers, tanks, agitators, pumps)	876,000	3.3
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack)	4,248,000	15.9
Absorption (4 tray towers including humidification chambers and mist eliminators, recirculation tanks, agitators, and pumps)	9,206,000	34.4
Stack gas reheat (4 indirect steam reheaters)	1,282,000	4.8
Reaction (tanks and agitators)	357,000	1.3
Solids separation (thickener, drum filters, tanks, agitators, pumps, and conveyor)	2,352,000	8.8
Solids disposal (onsite disposal facilities including reslurry tank, agitator, slurry disposal pumps, and pond water return pumps)	1,247,000	4.7
Subtotal	21,235,000	79.4
Services, utilities, and miscellaneous	1,274,000	4.8
Total process areas excluding pond construction	22,509,000	84.2
Pond construction	4,241,000	15.8
Total direct investment	26,750,000	100.0
Indirect Investment		
Engineering design and supervision	1,444,000	5.4
Architect and engineering contractor	321,000	1.2
Construction expense	3,746,000	14.0
Contractor's fees	1,167,000	4.4
Total indirect investment	6,688,000	25.0
Contingency	6,688,000	25.0
Total fixed investment	40,126,000	150.0
Other Capital Charges		
Allowance for startup and modifications	3,589,000	13.4
Interest during construction	4,815,000	18.0
Total depreciable investment	48,530,000	181.4
Land	837,000	3.1
Working capital	1,184,000	4.4
Total capital investment	50,551,000	188.9

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a. **Basis:**
Midwest plant location represents project beginning mid-1977, ending mid-1980.
Average cost basis for scaling, mid-1979.
Stack gas reheat to 175°F by indirect steam reheat.
Minimum in-process storage; only pumps are spared.
Disposal pond located 1 mi from power plant.
Investment requirements for flyash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 3. CITRATE PROCESS

SUMMARY OF ESTIMATED FIXED INVESTMENT[a]

(500-MW new coal-fired power unit, 3.5% S in coal; 1.2 lb SO_2/MBtu heat input allowable emission)

	Investment, $	Percent of total direct investment
Direct Investment		
Materials handling (unloading conveyor, elevator conveyor, pneumatic conveyor, feed storage bins)	804,000	2.1
Feed preparation (conveyors, tanks, agitators, pumps, feeders)	118,000	0.3
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack)	4,368,000	11.2
SO_2 absorption (4 packed tower absorbers including humidification chambers and mist eliminators, surge tanks, centrifugal pumps, compressor, strippers)	14,223,000	36.6
Stack gas reheat (4 indirect steam reheaters)	1,294,000	3.3
Chloride purge (feeder, tank, agitator, pump)	83,000	0.2
SO_2 reduction (reactor tanks, aging tanks, agitators, centrifugal pumps)	1,303,000	3.4
S separation and removal (flotation tanks, rotary drum filter, pumps, slurry tank, heat exchanger, settling tank, heaters, flash drum)	2,118,000	5.5
S storage and shipping (S receiving pit, heaters, S pump and storage tank)	814,000	2.1
Sulfate removal (coolers, agitators, centrifuge, tanks, pumps, and refrigeration)	985,000	2.5
H_2S generation (battery limit plant)	5,850,000	15.1
H_2 generation (battery limit plant)	4,680,000	12.0
Subtotal	36,640,000	94.3
Services, utilities, and miscellaneous	2,198,000	5.7
Total direct investment	38,838,000	100.0
Indirect Investment		
Engineering design and supervision	3,273,000	8.4
Architect and engineering contractor	818,000	2.1
Construction expense	5,208,000	13.4
Contractor's fees	1,548,000	4.0
Total indirect investment	10,847,000	27.9
Contingency	9,937,000	25.6
Total fixed investment	59,622,000	153.5
Other Capital Charges		
Allowance for startup and modifications	5,962,000	15.4
Interest during construction	7,155,000	18.4
Total depreciable investment	72,739,000	187.3
Land	39,000	0.1
Working capital	2,111,000	5.4
Total capital investment	74,889,000	192.8

[a] Basis:
Midwest plant location represents project beginning mid-1977, ending mid-1980.
Average cost basis for scaling, mid-1979.
Stack gas reheat to 175°F by indirect steam reheat.
Minimum in-process storage; only pumps are spared.
Investment requirements for flyash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.
Construction labor shortages with accompanying overtime pay incentive not considered.

Table 4. LIMESTONE SLURRY PROCESS

TOTAL AVERAGE ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, $	Percent of total annual revenue requirements
Direct Costs				
Raw materials				
Limestone	158,300 tons	7.00/ton	1,108,100	7.86
Total raw materials cost			1,108,100	7.86
Conversion costs				
Operating labor and supervision	25,990 man-hr	12.50/man-hr	324,900	2.30
Utilities				
Steam	489,300 MBtu	2.00/MBtu	978,600	6.94
Process water	247,400 kgal	0.12/kgal	29,700	0.21
Electricity	56,670,000 kWh	0.029/kWh	1,643,400	11.66
Maintenance				
Labor and material			1,822,800	12.93
Analyses	3,760 man-hr	17.00/man-hr	63,900	0.45
Total conversion costs			4,863,300	34.49
Total direct costs			5,971,400	42.35
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			2,800,600	19.86
Average cost of capital and taxes at 8.6% of total capital investment			4,190,600	29.72
Overheads				
Plant, 50% of conversion costs less utilities			1,105,800	7.84
Administrative, 10% of operating labor			32,500	0.23
Total indirect costs			8,129,500	57.65
Total annual revenue requirements			14,100,900	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/short ton S removed
Equivalent unit revenue requirements	4.03	9.40	0.45	408.01

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a. Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175° F.
S removed, 34,560 short ton/yr; solids disposal 192,000 tons/yr calcium solids including only hydrate water.
Investment and revenue requirement for removal and disposal of flyash excluded.
Total direct investment, $26,000,000; total depreciable investment, $46,677,000; and total capital investment, $48,728,000.

Table 5. GENERIC DOUBLE-ALKALI PROCESS

TOTAL AVERAGE ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, $	Total annual revenue requirement
Direct Costs				
Raw materials				
Lime	63,600 tons	42.00/ton	2,671,200	18.20
Soda ash	6,060 tons	90.00/ton	545,400	3.72
Total raw materials cost			3,216,600	21.92
Conversion costs				
Operating labor and supervision	34,500 man-hr	12.50/man-hr	431,300	2.94
Utilities				
Steam	489,300 MBtu	2.00/MBtu	978,600	6.67
Process water	241,500 kgal	0.12/kgal	29,000	0.20
Electricity	29,100,000 kWh	0.029/kWh	843,900	5.75
Maintenance				
Labor and material			1,027,600	7.00
Analyses	4,560 man-hr	17.00/man-hr	77,500	0.53
Total conversion costs			3,387,900	23.09
Total direct costs			6,604,500	45.01
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6% of total depreciable investment			2,911,800	19.84
Average cost of capital and taxes at 8.6% of total capital invesment			4,347,400	29.63
Overheads				
Plant, 50% of conversion costs less utilities			768,200	5.23
Administrative, 10% of operating labor			43,100	0.29
Total indirect costs			8,070,500	54.99
Total annual revenue requirements			14,675,000	100.00

	mills/kWh	$/ton coal burned	$/MBtu heat input	$/short ton S removed
Equivalent unit revenue requirements	4.19	9.78	0.47	424.62

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a. Basis:
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
S removed, 34,560 short ton/yr; solids disposal 144,690 tons/yr calcium solids including only hydrate water.
Investment and revenue requirement for removal and disposal of flyash excluded.
Total direct investment, $26,750,000; total depreciable investment, $48,530,000; and total capital investment, $50,551,000.

Table 6. CITRATE PROCESS

TOTAL AVERAGE ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission)

	Annual quantity	Unit cost, $	Total annual cost, $	Percent of total annual revenue requirements
Direct Costs				
Raw materials				
Lime	2,870 tons	42.00/ton	120,500	0.52
Soda ash	2,630 tons	90.00/ton	236,700	1.02
Citric acid	230 tons	1,340.00/ton	308,200	1.32
Natural gas	1,050,000 kft^3	3.50/kft^3	3,675,000	15.80
Catalyst			21,000	0.09
Total raw materials cost			4,361,400	18.75
Conversion costs				
Operating labor and supervision	67,920 man-hr	12.50/man-hr	849,000	3.65
Utilities				
Steam	1,027,500 MBtu	2.00/MBtu	2,055,000	8.83
Process water	2,492,500 kgal	0.06/kgal	149,600	0.64
Electricity	68,530,000 kWh	0.029/kWh	1,987,400	8.54
Maintenance				
Labor and material			2,330,300	10.03
Analyses	10,600 man-hr	17.00/man-hr	180,200	0.77
Total conversion costs			7,551,500	32.46
Total direct costs			11,912,900	51.21
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6% of total depreciable investment			4,364,300	18.76
Average cost of capital and taxes at 8.6% of total capital investment			6,440,500	27.69
Overheads				
Plant, 50% of conversion costs less utilities			1,679,800	7.22
Administrative, 10% of operating labor			84,900	0.36
Marketing, 10% of sales revenue			135,400	0.58
Total indirect costs			12,704,900	54.61
Gross annual revenue requirements			24,617,800	105.82
Byproduct Sales Revenue				
Sulfur	33,840 short tons	40.00/short ton	(1,353,600)	(5.82)
Total annual revenue requirements			23,264,200	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/short ton S removed	$/short ton S recovered
Equivalent unit revenue requirements	6.65	15.51	0.74	673.15	687.48

a. Basis:
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
S removed, 34,560 short ton /yr.
Investment and revenue requirement for removal and disposal of flyash excluded.
Total direct investment, $38,838,000; total depreciable investment, $72,739,000; and total capital investment, $74,889,000.

SHAWNEE LIMESTONE-LIME COMPUTER PROGRAM

In conjunction with the EPA-sponsored Shawnee test program, Bechtel and TVA have jointly developed a computer program capable of projecting comparative investment and revenue requirements for lime and limestone scrubbing systems. The computer program was developed to permit the estimation of relative economics of lime-limestone scrubbing systems for variations in process design alternatives (i.e., limestone vs lime scrubbing, alternative scrubber types, or alternative sludge disposal methods) or variations in the values of independent design variables (i.e., scrubber gas velocity and L/G ratio, alkali stoichiometry, slurry residence time, reheat temperature, and specific sludge disposal design). Although the program is not intended to compute the economics of an individual system to a high degree of accuracy, it is based on sufficient detail to allow the quick projection of preliminary conceptual design and costs for various limestone-lime case variations on a common design and cost basis.

The responsibility in the development of the computer program was shared by Bechtel and TVA. Bechtel's major responsibility was to analyze the results of the Shawnee scrubbing tests and develop models for calculating the overall material balance flow rates, stream compositions, and equipment size or capacity for the major process equipment. Bechtel provided TVA with a complete computer program for specifying this information. TVA was responsible for determining the size limitations of the required equipment for establishing the minimum number of parallel equipment trains, accumulating cost data for the major equipment items, and developing models for projecting equipment and field material costs as a function of equipment capacity. Utilizing these relationships TVA developed models to project the overall investment cost breakdown and a procedure for using the output of the material balance and investment models as input to a previously developed TVA program for projecting annual and lifetime revenue requirements.

Program Scope

The present computer program has the capability of projecting economics for either limestone or lime scrubbing utilizing a Turbulent Contact Absorber (TCA). The following alternative sludge disposal options may be evaluated.

1. Onsite pond
2. Thickener-onsite pond
3. Thickener-fixation (fee)
4. Thickener-filter-fixation (fee)

The following alternative operating profiles are presently available for projecting lifetime revenue requirements.

1. Profile (see below) similar to that utilized in Detailed Cost Estimates for Advanced Effluent Desulfurization Processes.

Operating Profile Option 1 Basis

Operating yr (plant age)	Annual capacity factor, % of nameplate rating
1-10	80
11-15	57
16-20	40
21-30	17
Average for 30-yr life	48.5

2. Historical power plant operating profile based on FPC Form 67 data and represented by the generalized equations given below:

Operating Profile Option 2 Basis

Operating yr (plant age)	Annual capacity factor (CF), % of nameplate rating (equation)
1-10	CF = 50 + 1.5 x plant age
11-15	CF = 65
16-30	CF = 92 - 1.8 x plant age
Average for 30-yr life	55.6

3. Variable profile with annual load factors as input.

Incorporation of a venturi-spray tower option is currently underway. Upon completion, it is planned to add spray tower and series scrubber options along with a water balance model relating makeup water requirements to rainfall and losses in the pond from seepage and evaporation.

Process Definition

The computer program was developed to model either limestone or lime scrubbing schemes. A flow-control diagram of the limestone scrubbing scheme modeled is shown in Figure 4. A plan and elevation diagram of the TCA scrubber arrangement is shown in Figure 5, and the pond construction diagram is shown in Figure 6. The absorbent preparation area for the lime scrubbing option is illustrated in Figure 7. The following is a general description of the limestone and lime processes.

Flue gas handling. Flue gas from the power unit ducts is fed to a common plenum, from which any number of scrubbing trains can be fed. To minimize the problems associated with gas distribution for such a system, separate fans are included on each side of the plenum. The power plant fans are conventional ID fans for balanced-draft boilers. The forced-draft scrubber fans are designed to over-

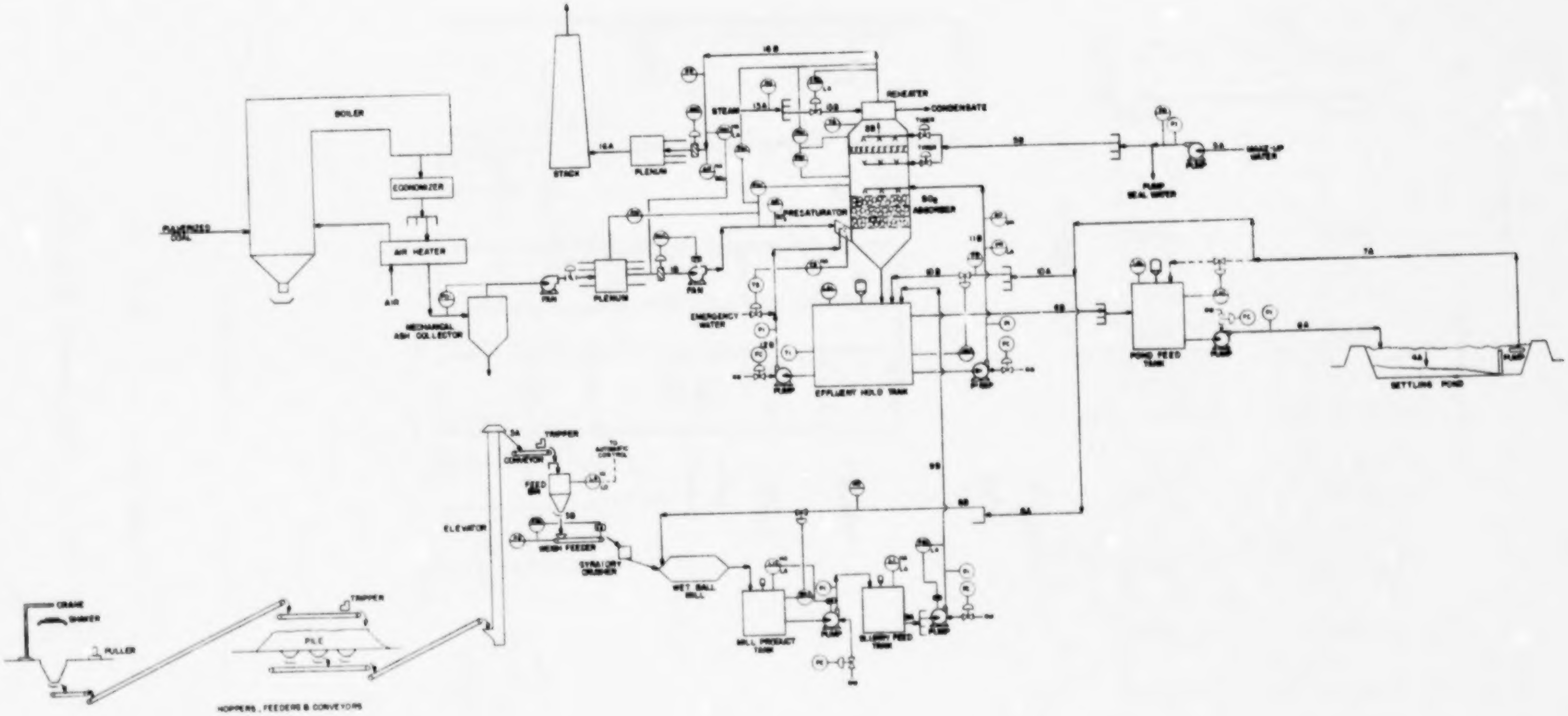
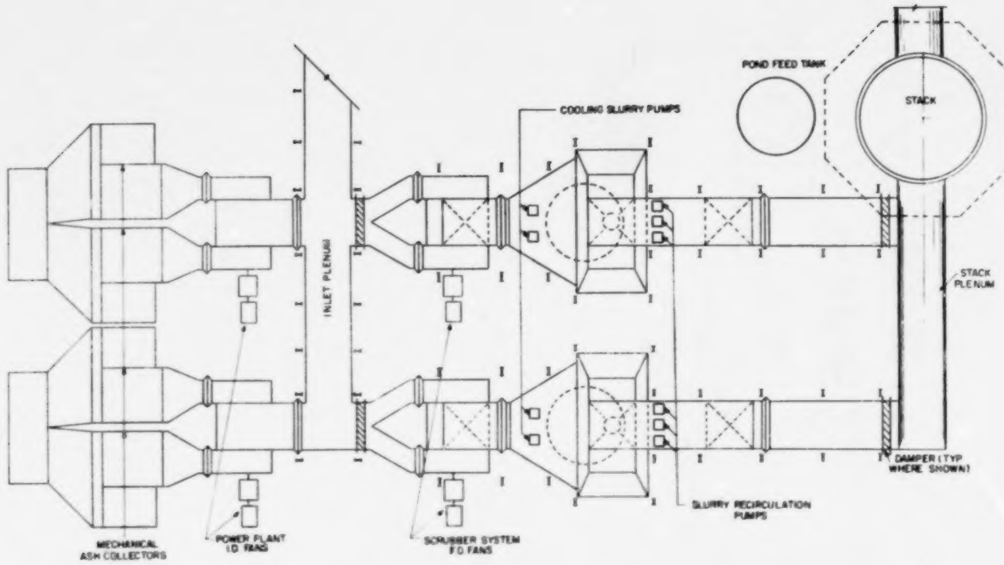
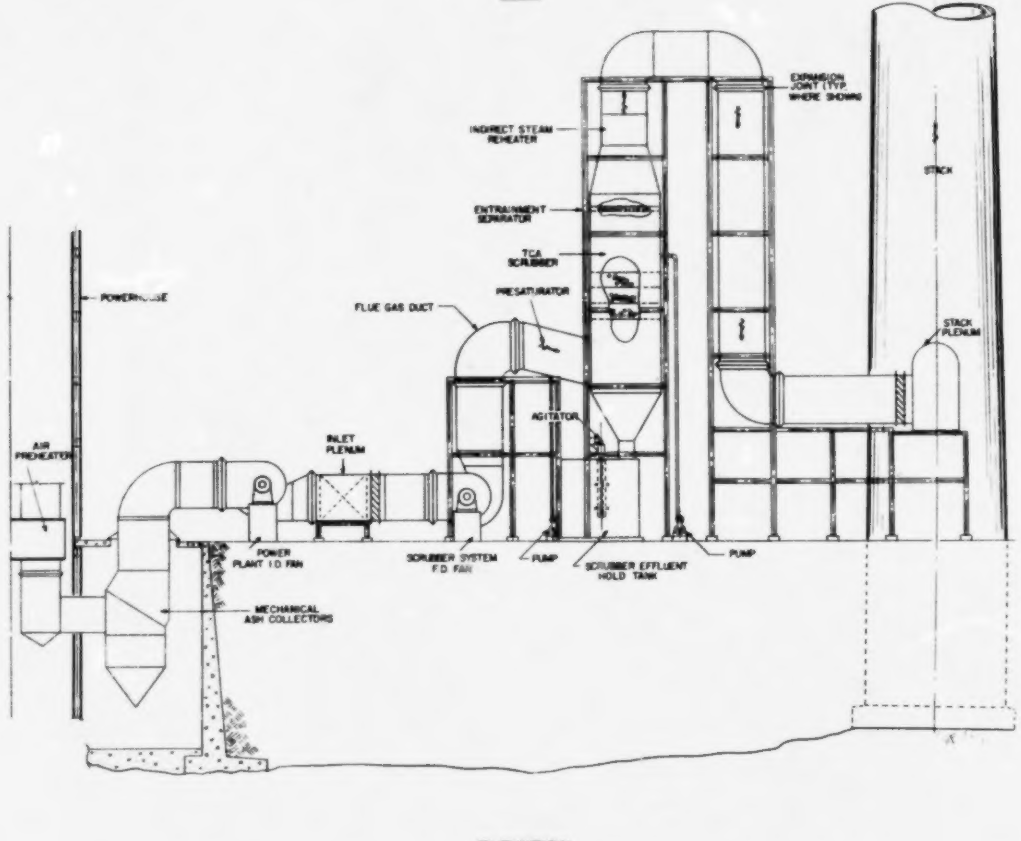


Figure 4. Limestone slurry process flow-control diagram.

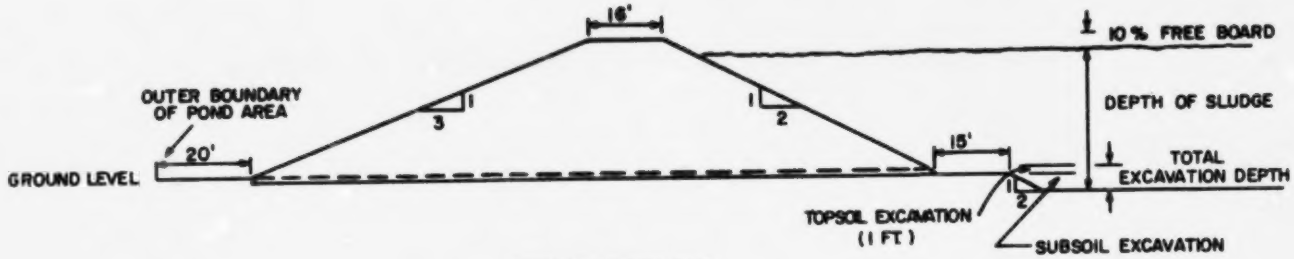


PLAN



ELEVATION

Figure 5. TCA scrubber arrangement diagram.



POND PERIMETER DIKE

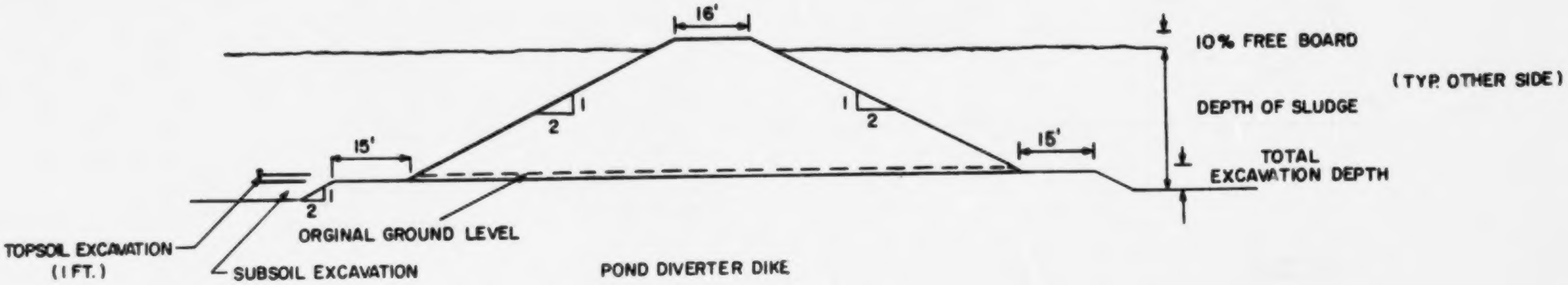


Figure 6. Pond construction diagram.

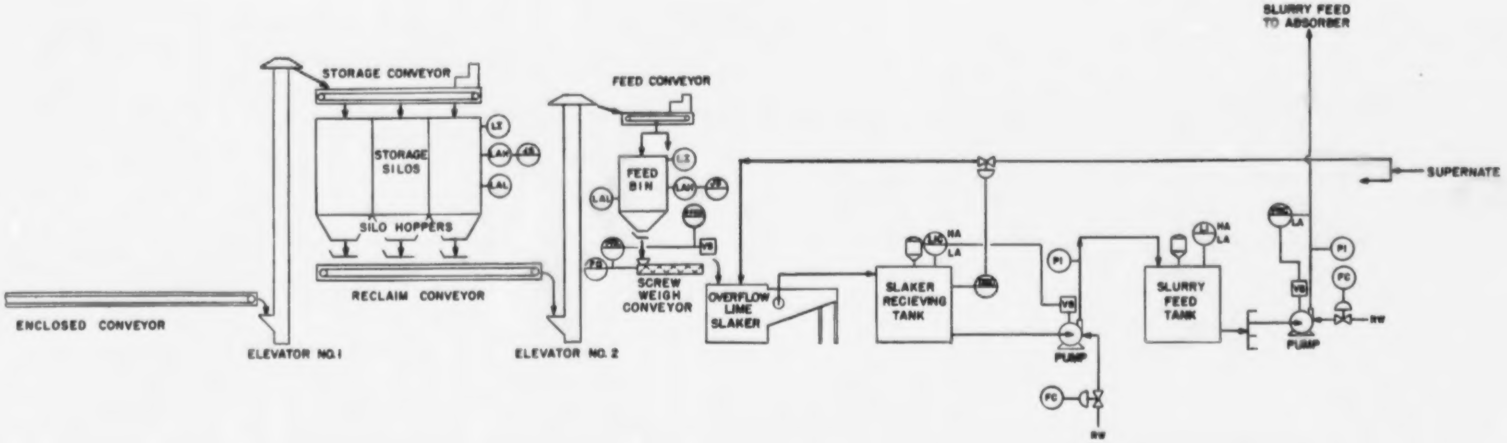


Figure 7. Lime slurry option absorbent preparation area.

come the pressure drop of the pollution control facilities. Since a dry fan system was selected, mechanical collectors are included in the process for removal of the coarse ash particles from the gas to protect the fans.

Flue gas from the scrubber fans is contacted with either a limestone or a lime slurry in TCA's equipped with a chevron-vane mist elimination system designed for upstream and downstream wash with fresh makeup water. The outlet gas from the scrubber is reheated to the desired temperature by indirect steam reheat and discharged to the stack plenum.

Raw material handling and preparation. For the limestone slurry process option the raw material handling and preparation area includes equipment for receiving limestone by truck or rail, a storage stockpile, live in-process limestone storage equipment, and the equipment for crushing and wet grinding the raw limestone to the desired size for feed to the scrubbers. The product slurry from the mills is pumped to a slurry feed tank adjacent to the scrubbing area for distribution to the scrubbers.

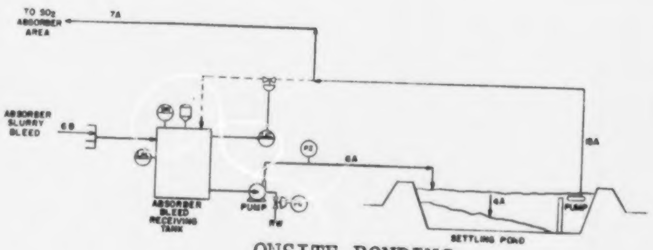
For the lime slurry process option the raw material handling and preparation area includes equipment for conveying lime from a calcining plant located 1500 ft from the area to a storage silo and slaking the lime. The product slurry from each of the slakers overflows to a slurry receiving tank from which it is pumped to a common slurry feed tank. The slurry is then pumped to the scrubbing area for distribution to the scrubbers.

Slurry recirculation loop. Makeup limestone or lime slurry from the slurry feed tank and recycled supernate from the sludge disposal system are fed to the absorber effluent hold tank for blending with the slurry draining from the absorber. The slurry is then recirculated to the absorber. A separate slurry stream is recirculated to the presaturator for partial humidification and resultant cooling of the gas upstream of the scrubber to insure protection of the scrubber lining and internals from the hot flue gas. The net quantity of reaction products formed in each scrubber overflows to a common absorber bleed receiving tank and is pumped to the waste disposal system.

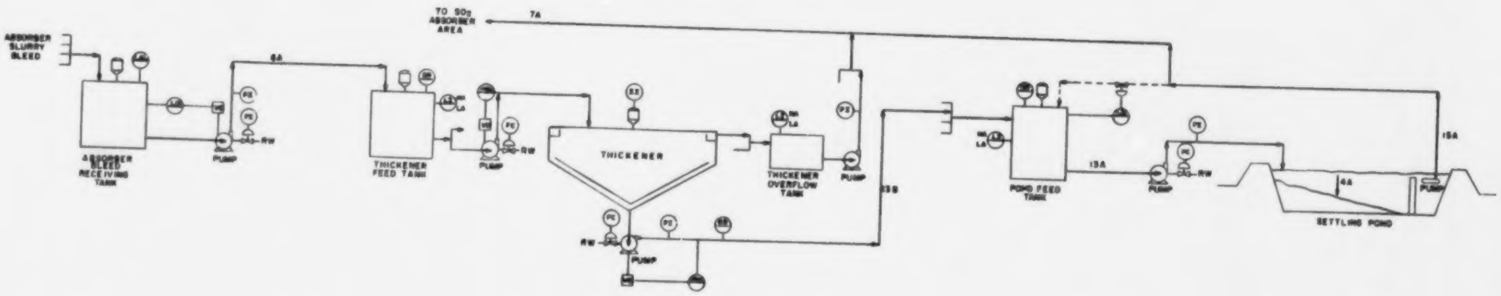
Waste disposal system. The present program utilizes the following four waste disposal system alternatives which are illustrated in Figures 8 and 9.

1. Onsite ponding. The slurry bleed stream is pumped from the absorber bleed receiving tank to an onsite pond where it is allowed to settle. Pond supernate is recycled to the wet ball mills and to the absorber effluent hold tanks to maintain a closed-loop process.

2. Thickener-ponding. The slurry bleed stream is thickened before being pumped to an onsite pond for further settling. Pond supernate and thickener overflow are recycled to the wet ball mills and to the absorber effluent hold tanks to maintain a closed-loop process.

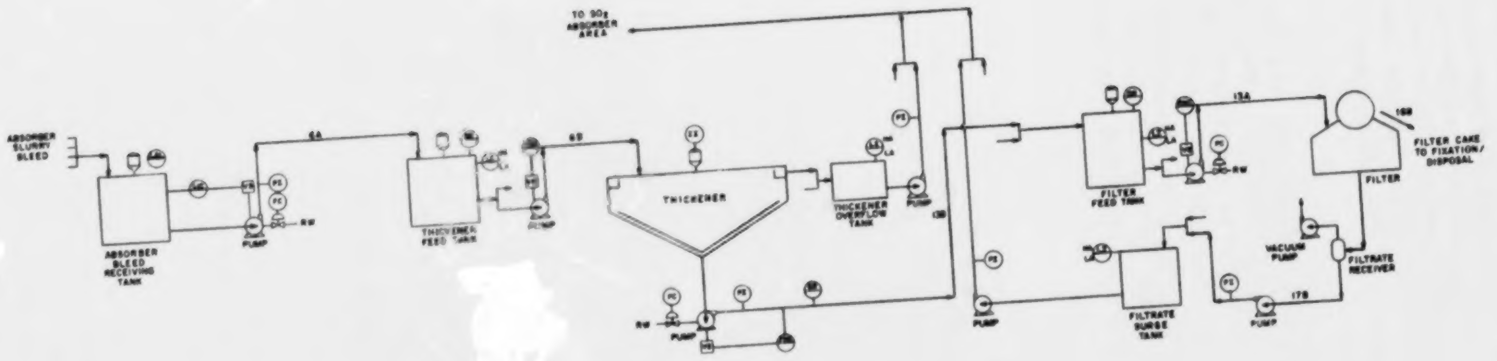


ONSITE PONDING

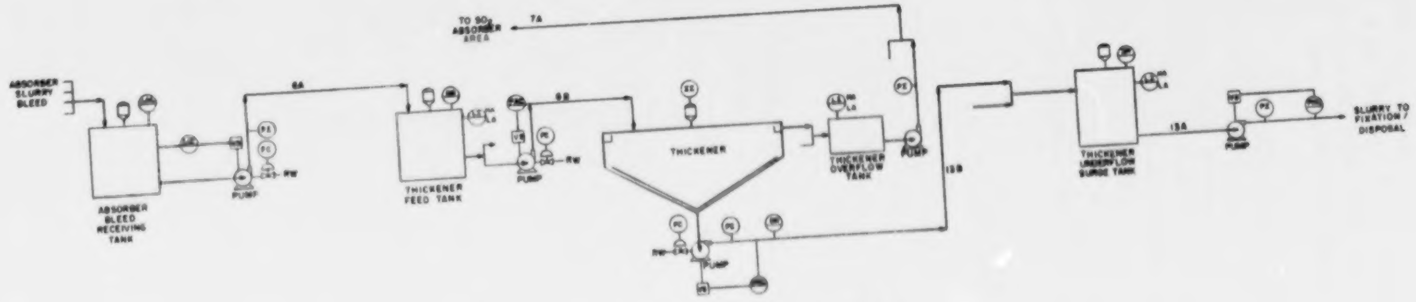


THICKENER--PONDING

Figure 8. Onsite pond disposal alternatives.



THICKENER-FILTER-FIXATION (FEE)



THICKENER-FIXATION (FEE)

Figure 9. Fixation alternatives.

3. Thickener-fixation (fee). The slurry bleed stream is first thickened. The thickener underflow is assumed to be chemically "fixed" followed by disposal as landfill. Facilities, and both investment and operating costs for thickening are included in the process design and the projected economics. The costs for chemical fixation and disposal are an input to the program as \$/ton of dry solids to be fixed and must include charges to account for the depreciation, capital charges, and operating expense for the fixation equipment and the disposal site.

4. Thickener-filter-fixation (fee). Thickeners and filters are utilized to dewater the slurry bleed stream in preparation for fixation. Facilities and both investment and operating costs for thickening and filtering the sludge are included in the process design and the projected economics. Costs for chemical fixation and disposal are an input to the program as \$/ton of dry solids to be fixed, and must include charges to account for the depreciation, capital charges, and operating expense for the fixation equipment and the disposal site.

Major Equipment Specifications

The following procedures are utilized for determining the size or specifications of the major process equipment in the limestone slurry process raw material handling and preparation area.

Gyratory crushers. Two parallel 50% capacity gyratory-type crushers are utilized to reduce the inlet stone size from minus 1-1/2 in. to minus 3/4 in. for feed to the ball mills.

Ball mills. The grinding mills are rubber-lined, open-circuit, overflow wet ball mills that have a 30% ball charge and produce a 60% slurry. The number of ball mills is determined by total mill hp calculated from the limestone throughput rate specified in the material balance and the fineness of grind and limestone hardness factors which are program inputs. The fineness of grind index factor is related to the desired particle size distribution of the ground limestone. One-mill systems are used for hp less than 200 and two parallel mill systems for hp between 200 and 5000. For hp greater than 5000, the number of parallel mill systems is determined assuming a maximum mill size of 2500 hp.

The lime slurry process raw material handling and preparation area utilizes the following major process equipment.

Lime conveyor. One enclosed belt conveyor is used to transport lime 1500 ft from an across-the-fence lime calcination plant to the storage silos. The conveyor operates at belt speeds of 100-300 ft/min and handles up to 75 tons/hr of lime.

Lime storage silo. A 15-day dead storage capacity is used to calculate the volume of the lime storage silos. The silos are concrete with the height of the actual storage section of the silo assumed to be one and a half times the diameter. Total height of the silo is equal to the height of the actual storage section plus the height of the carbon steel hopper plus 5 ft. Parallel storage silos are used for storage volumes greater than the capacity of the largest silo (147,200 ft^3).

Lime slaker. Lime is slaked at slurry concentrations of 20-25% solids in dual compartment, overflow slakers which can be designed with slaking capacities of up to 33 tons/hr . However, parallel slaking trains are used for lime capacities greater than 2 tons/hr. The number and size of parallel slakers required are determined based on the capacity of the largest size slaker available (33 tons/hr).

The following procedures are utilized for determining the size or specifications of the major process equipment in the scrubbing area.

Scrubbing trains. The number of parallel scrubbing trains is either an input to the program or is established as an override to the input value based on the minimum number of scrubber trains required. The minimum number of trains is calculated considering the saturated flue gas velocity and volumetric flow rate at the scrubber outlet in conjunction with the maximum cross-sectional area assumed for the TCA scrubber (1008 ft^2). Flue gas and slurry recirculation rates per train are calculated by dividing the total flow rates from the overall material balance model by the number of operating scrubbing trains.

Scrubbers. Scrubber cross-sectional area is calculated considering the outlet flue gas rate per train in conjunction with the specified scrubber design gas velocity. Number of scrubber grids, beds, and height of spheres per bed are inputs to the program. The height of the scrubber is assumed to remain constant for all scrubber sizes and internal configurations. A presaturator compartment is included at the scrubber inlet and chevron-type mist eliminators near the outlet. Materials of construction for the scrubbers and internals are listed below.

Shell:	rubber-lined carbon steel
Grids:	type 316L stainless steel
Spheres:	thermoplastic rubber (TPR)

Mist eliminator, slurry header, and nozzles: type 316L stainless steel

Reheaters. Reheater cross-sectional area is calculated based on the superficial gas velocity which is input to the program and the volumetric gasflow rate per train at scrubber outlet conditions. Reheater surface area requirements are calculated in two steps (1) surface area requirements for reheat to 150°F and (2) requirements for reheat to the specified reheat temperature. The portion of the reheater tubes required to reheat to 150°F are Inconel and the remaining tubes are Cor-Ten. Reheater design and costs are based on use of 1-in. tubes on a 2-in. square pitch.

Fans. The fans are centrifugal (double width, double inlet) with radial impellers and are constructed of carbon steel. They are equipped with variable-speed fluid drives. Fan hp is calculated based on the inlet gasflow rate per train and the calculated pressure drop for the scrubber, mist eliminator, reheater, and duct.

The following procedures are utilized for determining the size or specifications of the major process equipment for the various waste disposal area alternatives.

Thickeners. The thickeners are constructed of flake glass-lined carbon steel walls with 1-ft-thick concrete conical basins and are equipped with rake mechanisms. A concrete underflow tunnel including pumps and piping for transferring the slurry is included. Total thickener cross-sectional area is calculated by the material balance portion of the model as a function of the settling rate and settled solids density which are inputs into the program and the quantity of sludge in the effluent slurry calculated in the material balance model. Number of thickeners required is determined assuming a maximum thickener diameter of 400 ft. Thickener height is calculated as a function of the diameter.

Filters. Rotary drum filters constructed of carbon steel and equipped with a vacuum pump, a filtrate pump, and a vacuum receiver are utilized. Filter size is presently determined as a function of the filtration rate, which is a program input, expressed in tons of dry solids/ft^2/day in conjunction with the total quantity of sludge. The minimum and maximum size filters considered have effective filtration areas of 50 and 900 ft^2, respectively. Single filters are used up to required filtration areas of 100 ft^2. For total filtration areas between 100 and 1800 ft^2, two parallel filters are assumed. For total filtration areas greater than 1800 ft^2, the number and size of parallel filters required are determined based on the capacity of the largest filter size.

The size or specifications of tanks, agitators, and pumps for each of the areas are determined by utilizing the following procedures.

Tanks. Tank volume is calculated based on the residence time which is either a program input or assumed with an additional 10% volume added for freeboard. All tanks are constructed of carbon steel and the slurry tanks are flake glass lined. Except for the absorber bleed receiving tanks and the thickener overflow tanks, each tank is designed with diameter equal to height up to a maximum height of 60 ft. For tanks larger than 60-ft diameter, tank height is fixed at 60 ft and diameter is calculated. Absorber bleed receiving tank height is equal to the effluent hold tank height and the diameter is calculated. Thickener overflow tank height is set equal to the height of the thickener and the diameter calculated. As an override to the calculated diameter a minimum diameter equal to one-half the height is fixed for all tanks. The thickener and filter feed tanks are not used unless more than one thickener or filter is required.

Agitators. All slurry tanks are equipped with a four-blade, pitched blade, turbine agitator. Agitator hp requirements are calculated on the basis of total torque which is a function of the degree of agitation required (expressed as torque/unit volume), total tank volume, tank diameter, and the slurry specific gravity. Unit torque (torque/unit volume) for each tank is determined as a function of the percent solids in the slurry.

Slurry pumps. All slurry pumps are rubber lined, centrifugal with water seals, and are equipped with either a variable- or constant-speed drive. Pumps are usually spared with number of operating pumps determined by the maximum available pump size of 20,000 gpm.

Water pumps. Vertical, multiple-stage, turbine makeup water pumps capable of providing a static head of 200 ft are provided for each 10,000 gpm of water required. The pumps are carbon steel and spared.

Field Material Specifications

Costs for field materials are based on the materials of construction or specifications discussed below.

Piping. Carbon steel pipe and gate valves are used for all waterlines including pond supernate. For slurry lines, stainless steel pipe is used for lines less than 3 in. diameter, whereas for all larger size lines rubber-lined carbon steel piping is used. Stainless steel strainers are used for pipes less than 4 in. diameter and rubber-lined strainers are used for 4-in.-diameter and larger pipes.

For slurry pipes less than 3 in. diameter, stainless steel plug valves are used. Eccentric plug valves are used for slurry pipe between 3 and 20 in. diameter, and knife gate valves are used for pipes greater than 20 in. diameter. Handwheel operators are used for valves less than 12 in. diameter and air cylinder actuators for larger valves. Typical piping layouts are assumed as functions of flow capacities and number of trains and costs are correlated to flow rates in gpm. Control valve costs are included in instrumentation. Costs are included for a rubber-lined downcomer from the TCA scrubber to the effluent hold tank and a spare slurry disposal line to the pond.

Ductwork. Costs are included for the inlet plenum and all ductwork between the inlet and stack plenums including insulation. Costs for the stack plenum are not included since it is also required for power plants without FGD systems. Stack plenum elevation is set equal to effluent hold tank heights with a minimum elevation of 20 ft for small hold tanks. Each scrubber train includes two dampers with costs included for expansion joints.

Materials of construction for all ductwork is Cor-Ten with the exception of the ductwork between the scrubber and reheater outlet which is type 316 stainless steel. Duct size is based on a square cross section and a nominal design velocity of 3000 ft/min at inlet conditions (approximately 300°F). Duct cross-sectional area downstream of the scrubbers and reheaters is set equal to the upstream cross-sectional area.

Foundations. Concrete foundations for each equipment item are fixed according to equipment sizes. Foundations for the structure are estimated on the basis of the weight of the structure.

Pond construction. Disposal pond size is calculated based on a square configuration with a diverter dike three-fourths the length of one side. A pond construction diagram is shown in Figure 6.

A separate model is included to design and cost the onsite pond and the following design information is displayed.

Sludge to be disposed of over life of plant in yd^3
Equivalent disposal volume in acre-ft
Pond depth in ft
Excavation required in construction of pond in yd^3

Pond size in acres
Breakdown of surface area for the dike and pond

The pond model is based on either unlined, clay-lined, or synthetic-lined design and includes the following options in running the program.

Fixed-depth pond
Optimum-depth pond based on minimum pond investment
Optimum-depth pond based on minimum pond investment with available acreage and maximum excavation depth as overriding constraints

In addition to specifying pond design, the model also itemizes the breakdown of projected pond costs.

Structural. Structural estimates are based upon the structure arrangement shown in Figure 5. The total quantity of structure required and the corresponding costs are related to effluent hold tank volume, scrubber cross-sectional area, and number of scrubbing trains.

Electrical. The electrical estimate is divided into four sections: (1) costs of feeder cables from the power plant transformer yard to field modules for each area, (2) transformer costs for each area, (3) costs of power supply from area field modules to individual motors, and (4) motor control costs between remote control center, field module location, and individual motors for each area. For each area, total connected motor hp is calculated for use in establishing costs for (1) and (2). Costs for (3) and (4) are based on individual motor sizes and number of connected motors. A typical layout is assumed for each area in reference to the power plant transformer yard, remote control center, and other areas.

Instrumentation. A typical control diagram for the limestone slurry process is shown in Figure 4. Control diagrams for the lime preparation area and the alternate sludge disposal options are shown in Figures 8 and 9. Instrumentation costs are based on (1) fixed costs for instruments which do not change in size and cost with equipment and pipe size variations and (2) variable costs for instruments which increase in size and cost as equipment and pipe sizes increase. Each of these costs may be dependent upon number of scrubbing trains, number of ball mills, number of pumps, etc. Costs are included for control valves, graphic and panelboards, annunciator, air dryers and piping, and instrument cable and wiring systems.

Buildings. The control room and motor control center are integrated with the power plant and prorated costs are included. For the limestone process, costs are included for a building to house the grinding facilities. The building size is determined by number of ball mill systems and ball mill size (hp). The cost and size of a building to house the drum filters are determined by number of filters and filter size (effective filtration area).

Services and miscellaneous. Services and miscellaneous costs are based on the definition and costs given in Detailed Cost Estimates for Advanced Effluent

Desulfurization Processes scaled proportional to the relative power unit size ratio raised to the 0.6 power for all cases.

Program Inputs

The overall computer program for the options presently available requires a minimum of 13 lines of input. Additional input is required for the program option in which a variable operating profile is selected rather than one of the built-in profiles. A detailed list of the inputs to the program along with the format will be provided to users when the program is made available. To illustrate the requirements for running the program, however, a generalized list of the inputs is given below.

Boiler characteristics
  Megawatts
  Heat rate, Btu/kWh
  Excess air (including inleak), %
  Hot gas temperature, ^oF
  Coal analysis, wt % of C, H, O, N, S, Cl, ash, and H_2O
  Percent S overhead
  Percent ash overhead
  Heating value of coal, Btu/lb

Mechanical collector (or ESP) efficiency, %

Alkali feed
  Chemical composition and moisture content of limestone or lime
  Chemical composition (alkalinity) of flyash

Scrubber system variables
  Number of TCA beds and grids
  Height of spheres per bed, in.
  Scrubber gas velocity, ft/sec
  L/G ratio, gal/kft^3
  Limestone or lime stoichiometry, mols Ca/mol SO_2 absorbed
  Effluent hold tank residence time, min
  Percent SO_2 oxidized in system
  Solids in recirculated slurry, wt %
  Number of operating scrubbing trains
  Number of redundant scrubbing trains

Solids disposal system
  Solids in system sludge discharge (settled density), wt %
  Solids in thickener discharge, wt %
  Thickener solids settling rate, ft/hr

Steam reheater (inline)
  Saturated steam temperature, ^oF
  Heat of vaporization of steam, Btu/lb
  Outlet flue gas (stack gas) temperature, ^oF
  Superficial gas velocity (face velocity), ft/sec

Limestone properties (for determining grinding mill size)
  Limestone hardness work index factor
  Limestone fineness of grind index factor

Sludge disposal option selection code
  Onsite ponding
  Thickener-ponding
  Thickener-fixation (fee)
  Thickener-filter-fixation (fee)

Disposal pond
  Available land for construction of pond, acre
  Final depth of sludge in pond, ft
  Maximum excavation depth in construction of pond, ft
  Distance from scrubber area to pond, ft
  Pond lining selection
    1. Unlined
    2. Clay lined (specify depth)
    3. Synthetic lined

Thickener
  Sludge settling rate, ft/hr
  Concentration of solids at outlet of thickener, wt %

Filter
  Sludge filtration rate, tons dry sludge/ft^2/day
  Concentration of solids in filter cake, wt %

Capital investment
  Chemical Engineering material cost index
  Chemical Engineering labor cost index
  Basis of investment estimate, yr
  Land cost, $/acre
  Clay cost in \$/yd^3 or synthetic liner installation labor cost in \$/yd^2
  Indirect investment cost factors
    Engineering design and supervision, %
    Construction expenses, %
    Contractor fees, %
    Contingency, %
    Allowance for startup and modification, %
    Interest during construction, %

Annual and lifetime revenue requirement
  Operating schedule selection
  Cost of money, %
  Maintenance rate, %
    Direct investment
    Pond

Capital charges
  Cost of capital and taxes, % of undepreciated investment
  Insurance and interim replacements, % of fixed investment

Overhead factors
  Plant, % of conversion costs less utilities
  Administrative, % of operating labor and supervision

Unit costs for direct operating cost items
  Limestone, $/ton
  Lime, $/ton

Operating labor and supervision, \$/man-hr
Steam, \$/klb
Process water, \$/kgal
Electricity, \$/kWh
Analyses, \$/hr

Offsite sludge disposal, \$/ton dry sludge

Program Outputs

The outputs of the overall computer program include (1) a detailed material balance including properties of the major streams, (2) specifications of the scrubbing equipment, (3) a detailed breakdown of the projected capital investment requirements, (4) an itemized breakdown of the projected revenue requirements by component for the first year of operation of the system, and (5) a lifetime revenue requirement analyses showing projected costs for each year of operation of the plant, as well as lifetime cumulative and discounted costs and equivalent unit revenue requirements. The output for an example run of the program is shown at the end of the paper. Details concerning the output of the program can best be seen by reviewing this example.

Potential Use of the Program

The current program may be used to estimate the economics for limestone or lime scrubbing systems utilizing a TCA for any of the four sludge disposal options discussed earlier. The effect of variations in any of the inputs, such as scrubber gas velocity, degree of SO_2 removal, reheat temperature, alkali stoichiometry, L/G ratio, etc., may be determined.

The limestone and lime scrubbing options may be run with SO_2 removal, stoichiometry, and L/G ratio all specified. For the limestone option, SO_2 removal and stoichiometry may be specified and L/G ratio calculated or SO_2 removal and L/G ratio may be specified and stoichiometry calculated.

The lifetime operating profile options permit the program to be run for either limestone or lime scrubbing to compare the effect of operating profile variations on economics.

Upon completion of the overall effort, the program will be useful for projecting a complete conceptual design package for lime-limestone scrubbing including material balance, capital investment estimate, and projected revenue requirements. It is expected that the program will be used by utility companies and engineering contractors involved in the selection and design of SO_2 removal facilities for specific applications. It is not intended to be used for projecting a final design of a given system, but to assist in the evaluation of system alternatives prior to development of a detailed design. Also, the program will be useful for evaluating the potential impact of various process variables on economics as a guide for planning research and development activities.

Program Limitations

The program is designed to consider power unit sizes ranging from 100-1300 MW. The SO_2 removal models are valid for coals ranging in S contents from approximately 2-5% S. To limit the extremely wide variations in equipment sizes and layout configurations which can result with variations in key independent variables, a range of values for the following variables was established:

Scrubber gas velocity	8-12.5 ft/sec
Liquor recirculation rate, L/G ratio	25-75 gal/kft^3
Slurry residence time in hold tank	2-25 min

The minimum number of scrubbing trains for the smallest size unit was arbitrarily set equal to two. A maximum of nine scrubbing trains is required for the maximum power unit size (1300 MW), the minimum design gas velocity (8 ft/sec), and an assumed maximum TCA scrubber size of 1008 ft^2.

Because of the wide variations in process layout which are encountered for retrofit installations, the investment and revenue requirement projections are valid only for new units which can be designed with essentially similar layouts.

In addition to these limitations, there are a number of program inputs which impose additional restrictions to the validity of the program which are not as apparent. For example, the forced-through material balance option requires that SO_2 removal, stoichiometry, and L/G ratio all be specified. Unless typical values of these variables are known, it is possible that the specified values of stoichiometry and L/G ratio would not yield the input value for SO_2 removal efficiency or would result in unrealistic values of pH. As discussed above, other inputs to the program should be modeled as functions of the chemistry and operating conditions in the scrubber rather than input. It must be remembered that the program is composed of generalized design and cost models, therefore, the overall results predicted by the program should not be considered to be absolute.

Future Program Additions or Modifications

The following additions or modifications are presently planned for incorporation into the overall program.

1. A venturi-spray tower scrubbing system
2. A spray tower scrubbing system
3. A series scrubbing, high oxidation-high utilization scrubbing system
4. A water balance model to relate makeup water requirements to rainfall, and losses in the pond from seepage and evaporation

As sufficient data become available, models will be incorporated into the program to relate the following items to the operating conditions in the scrubber.

1. Sulfate saturation
2. Sulfite oxidation
3. Slurry settled density
4. Slurry settling rate
5. Particulate removal efficiency
6. Entrainment
7. Effect of magnesium ion on SO_2 removal
8. SO_2 removal efficiency

Program Availability

Since the current version of the program covers only a single scrubber type, it is not yet ready for widespread application. Plans are to wait until the present program scope has been completed and other scrubbing options are available before preparing a detailed users manual and releasing the program for general use. Arrangements for maintaining and distributing the overall program upon completion are being outlined by EPA and TVA. Until the program is complete, the current version will be made available to users on a limited basis only.

INPUT DATA FOR EXAMPLE PROGRAM RUN

```
 TYPE IN LINES 1-13 ACCORDING TO PROGRAM DESCRIPTION MANUAL
END OR NEXT



0 0 0 0 0 0 0 0 0 0
1 1 1 1 1
1 1 1 1 1 1 1 1 1 1 1 1 1 1 0 1 1
EXAMPLE OUTPUT LIMESTONE SCRUBBING WITH ONSITE PONDING
0 500 9000 10500 33 300 2 175 470 751
57.56 4.14 7.00 1.29 3.12 0.15 16.0 10.74 95 0.08
50 0 12.5 25 470 10 1 1.50 1 .15 0 4.85 5 0 0 15 40 .2 35 20 60 1.2
2 3 4 5
10 2.45 4 0 .1 264.9 237.9 1979 1980
1 0 9999 3500 25 25 5280 1 12 2.5
9 16 5 10 8 12
11.6 8 3 50 10 17.2 2.7
7.0 40.0 12.50 2.0 0.12 0.029 17
END
```

```
EXAMPLE OUTPUT LIMESTONE SCRUBBING WITH ONSITE PONDING

                    *** INPUTS ***

BOILER CHARACTERISTICS
------ ---------------

MEGAWATTS =    500.

BOILER HEAT RATE =  9000. BTU/KWH

EXCESS AIR =  33. PERCENT, INCLUDING LEAKAGE

HOT GAS TEMPERATURE = 300. DEG F


COAL ANALYSIS, WT % AS FIRED :

 C        H        O        N        S        CL      ASH     H2O
57.56    4.14     7.00     1.29     3.12     0.15    16.00   10.74

SULFUR OVERHEAD =  95.0 PERCENT

ASH OVERHEAD =    0.1 PERCENT

HEATING VALUE OF COAL = 10500. BTU/LB


MECHANICAL COLLECTOR (OR ELECTROSTATIC PRECIPITATOR)
---------- --------- --- ------------- -------------

NONE


ALKALI
------

LIMESTONE :

        CACO3          =  95.00 WT % DRY BASIS
        SOLUBLE MGO =    0.15
        INERTS         =   4.85
        MOISTURE CONTENT =   5.00 LB H2O/100 LBS DRY LIMESTONE
        LIMESTONE HARDNESS WORK INDEX FACTOR = 10.00
        LIMESTONE DEGREE OF GRIND FACTOR =  2.45

FLY ASH :

        SOLUBLE CAO =    0.0  WT %
        SOLUBLE MGO =    0.0
        INERTS         = 100.00
```

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```
SCRUBBER SYSTEM VARIABLES
-------- ------ ---------

NUMBER OF OPERATING SCRUBBING TRAINS =   4

NUMBER OF REDUNDANT SCRUBBING TRAINS =   0

NUMBER OF BEDS =   3

NUMBER OF GRIDS =   4

HEIGHT OF SPHERES PER BED =  5.0 INCHES

LIQUID-TO-GAS RATIO =  50. GAL/1000 ACFM

SCRUBBER GAS VELOCITY = 12.5 FT/SEC

SO2 CONCENTRATION IN SCRUBBER OUTLET GAS =  470. PPM

ENTRAINMENT LEVEL = 0.10 WT %

EHT RESIDENCE TIME =  10.0 MIN

SO2 OXIDIZED IN SYSTEM =  20.0 PERCENT

SOLIDS IN RECIRCULATED SLURRY =  15.0 WT %

SOLIDS DISPOSAL SYSTEM
------ -------- ------

SOLIDS IN SYSTEM SLUDGE DISCHARGE =  40.0 WT %

POND DEPTH = 25.00 FT

MAXIMUM EXCAVATION = 25.00 FT

DISTANCE TO POND = 5280. FT

POND LINED WITH 12.0INCHES CLAY

SOLIDS IN CLARIFIER DISCHARGE = 35.0 WT %

CLARIFIER SOLIDS SETTLING RATE = 0.20 FT/HR


STEAM REHEATER (IN-LINE)
----- -------- ---------

SATURATED STEAM TEMPERATURE =  470. DEG F

HEAT OF VAPORIZATION OF STEAM =  751. BTU/LB

OUTLET FLUE GAS TEMPERATURE = 175. DEG F

SUPERFICIAL GAS VELOCITY (FACE VELOCITY) =  25.0 FT/SEC
```

```
EXAMPLE OUTPUT LIMESTONE SCRUBBING WITH ONSITE PONDING


                    *** OUTPUTS ***


HOT GAS TO SCRUBBER
--- --- -- --------

          MOLE PERCENT      LB-MOLE/HR      LB/HR

CO2         12.315         0.2054E+05     0.9040E+06
HCL          0.011         0.1813E+02     0.6612E+03
SO2          0.238         0.3962E+03     0.2538E+05
O2           4.827         0.8050E+04     0.2576E+06
N2          73.867         0.1232E+06     0.3452E+07
H2O          8.743         0.1458E+05     0.2627E+06

SO2 CONCENTRATION IN SCRUBBER INLET GAS = 2376. PPM

FLY ASH =  0.01 GRAINS/SCF (WET)  OR      55. LB/HR

          SOLUBLE CAO IN FLY ASH =       0. LB/HR
          SOLUBLE MGO IN FLY ASH =       0.

HOT GAS FLOW RATE =  .1054E+07 SCFM (60 DEG F, 1 ATM)
                  =  .1540E+07 ACFM (300. DEG F, 1 ATM)

CORRESPONDING COAL FIRING RATE =  .4286E+06 LB/HR

HOT GAS HUMIDITY =  0.057 LB H2O/LB DRY GAS

WET BULB TEMPERATURE = 127. DEG F


WET GAS FROM SCRUBBER
--- --- ---- --------

          MOLE PERCENT      LB-MOLE/HR      LB/HR

CO2         11.682         0.2087E+05     0.9185E+06
SO2          0.047         0.8400E+02     0.5381E+04
O2           4.489         0.8019E+04     0.2566E+06
N2          68.962         0.1232E+06     0.3452E+07
H2O         14.820         0.2648E+05     0.4770E+06

SO2 CONCENTRATION IN SCRUBBER OUTLET GAS =  470. PPM

FLY ASH =  0.000 GRAINS/SCF (WET) OR     0. LB/HR

TOTAL WATER PICKUP =  439.  GPM
          INCLUDING   10.2 GPM ENTRAINMENT

WET GAS FLOW RATE =  .1128E+07 SCFM (60 DEG F, 1 ATM)
                  =  .1274E+07 ACFM (127. DEG F, 1 ATM)

WET GAS SATURATION HUMIDITY =  0.103 LB H2O/LB DRY GAS
```

```
FLUE GAS TO STACK
---- --- -- -----

          MOLE PERCENT       LB-MOLE/HR       LB/HR

CO2          11.664         0.2087E+05       0.9185E+06
SO2           0.047         0.8400E+02       0.5381E+04
O2            4.482         0.8019E+04       0.2566E+06
N2           68.853         0.1232E+06       0.3452E+07
H2O          14.955         0.2676E+05       0.4821E+06

SO2 CONCENTRATION IN STACK GAS =  469. PPM

FLY ASH =  0.000 GRAINS/SCF (WET) OR     0. LB/HR

STACK GAS FLOW RATE =  .1130E+07 SCFM (60 DEG F, 1 ATM)
                    =  .1380E+07 ACFM (175. DEG F, 1 ATM)


STEAM REHEATER (IN-LINE)
----- -------- ---------

SUPERFICIAL GAS VELOCITY (FACE VELOCITY) =  25.0 FT/SEC

SQUARE PIPE PITCH = 2 TIMES ACTUAL PIPE O.D.

SATURATED STEAM TEMPERATURE =  470. DEG F

OUTLET FLUE GAS TEMPERATURE = 175. DEG F

REQUIRED HEAT INPUT TO REHEATER = 0.6684E+08 BTU/HR

STEAM CONSUMPTION = 0.9166E+05 LBS/HR


                                    HEAT TRANSFER
OUTSIDE PIPE    PRESSURE DROP,     COEFFICIENT,
DIAMETER, IN.   IN. H2O            BTU/HR FT2 DEG F
  1.00              0.76             0.2082E+02


                REHEATER          NUMBER OF
                OUTSIDE PIPE      PIPES PER      NUMBER OF
                AREA, SQ FT       BANK PER       BANKS (ROWS)
                PER TRAIN         TRAIN          PER TRAIN

INCONEL         0.1285E+04            87              3
CORTEN          0.1313E+04            87              4
TOTAL           0.2597E+04            87              7
```

```
SCRUBBER SYSTEM
-------- ------

TOTAL NUMBER OF SCRUBBING TRAINS (OPERATING+REDUNDANT) =  4

SO2 REMOVAL =  78.8 PERCENT

PARTICULATE REMOVAL IN SCRUBBER SYSTEM =  99.50 PERCENT

PRESSURE DROP ACROSS  3 BEDS =  8.0 IN. H2O

LIMESTONE ADDITION = 0.4271E+05 LB/HR DRY LIMESTONE

LIMESTONE STOICHIOMETRY  =  1.30 MOLE CACO3 ADDED AS LIMESTONE
                                 PER MOLE SO2 ABSORBED

SOLUBLE CAO FROM FLY ASH =  0.0  MOLE PER MOLE SO2 ABSORBED

TOTAL SOLUBLE MGO        =  0.01 MOLE PER MOLE SO2 ABSORBED

TOTAL STOICHIOMETRY      =  1.30 MOLE SOLUBLE (CA+MG)
                                 PER MOLE SO2 ABSORBED

SCRUBBER INLET LIQUOR PH =  5.43

MAKE UP WATER =  602. GPM

 CROSS-SECTIONAL AREA PER SCRUBBER =   425. SQ FT


SOLIDS DISPOSAL SYSTEM
------ -------- ------

TOTAL CLARIFIER(S) CROSS-SECTIONAL AREA =  16271. SQ FT


SYSTEM SLUDGE DISCHARGE
------ ------ ---------
                                                      SOLID   LIQUID
                                                      COMP,   COMP,
SPECIES             LB-MOLE/HR     LB/HR              WT %    PPM

CASO3 .1/2 H2O      0.2496E+03     0.3223E+05         60.22
CASO4 .2H2O         0.6140E+02     0.1057E+05         19.74
CACO3               0.8592E+02     0.8601E+04         16.07
INSOLUBLES          ----------     0.2126E+04          3.97
H2O                 0.4393E+04     0.7915E+05
CA++                0.8586E+01     0.3441E+03                  4286.
MG++                0.1575E+01     0.3828E+02                   477.
SO3--               0.1352E+00     0.1083E+02                   135.
SO4--               0.1038E+01     0.9974E+02                  1242.
CL-                 0.1797E+02     0.6369E+03                  7934.

TOTAL DISCHARGE FLOW RATE = 0.1338E+06 LB/HR
                          =   202.     GPM

TOTAL DISSOLVED SOLIDS IN DISCHARGE LIQUID = 14074. PPM

DISCHARGE LIQUID PH =  7.27
```

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SCRUBBER SLURRY BLEED
-------- ------ -----

SPECIES              LB-MOLE/HR    LB/HR

CASO3 .1/2 H2O       0.2496E+03    0.3223E+05
CASO4 .2H2O          0.6140E+02    0.1057E+05
CACO3                0.8592E+02    0.8601E+04
INSOLUBLES           ----------    0.2126E+04
H2O                  0.1660E+05    0.2990E+06
CA++                 0.3244E+02    0.1300E+04
MG++                 0.5949E+01    0.1446E+03
SO3--                0.5109E+00    0.4090E+02
SO4--                0.3923E+01    0.3768E+03
CL-                  0.6787E+02    0.2406E+04

TOTAL FLOW RATE = 0.3568E+06 LB/HR
                =    648.    GPM


TOTAL SUPERNATE RETURN
----- --------- ------

SPECIES              LB-MOLE/HR    LB/HR

H2O                  0.1220E+05    0.2199E+06
CA++                 0.2385E+02    0.9559E+03
MG++                 0.4374E+01    0.1063E+03
SO3--                0.3757E+00    0.3007E+02
SO4--                0.2884E+01    0.2771E+03
CL-                  0.4991E+02    0.1769E+04

TOTAL FLOW RATE = 0.2230E+06 LB/HR
                =    446.    GPM


SUPERNATE TO WET BALL MILL
--------- -- --- ---- ----

SPECIES              LB-MOLE/HR    LB/HR

H2O                  0.1441E+04    0.2597E+05
CA++                 0.2817E+01    0.1129E+03
MG++                 0.5167E+00    0.1256E+02
SO3--                0.4437E-01    0.3552E+01
SO4--                0.3407E+00    0.3272E+02
CL-                  0.5895E+01    0.2090E+03

TOTAL FLOW RATE = 0.2634E+05 LB/HR
                =     53.    GPM
```

```
LIMESTONE SLURRY FEED
--------- ------ ----

SPECIES             LB-MOLE/HR   LB/HR

CACO3               0.4054E+03   0.4058E+05
SOLUBLE MGO         0.1509E+01   0.6407E+02
INSOLUBLES          ----------   0.2072E+04
H2O                 0.1558E+04   0.2807E+05
CA++                0.3045E+01   0.1221E+03
MG++                0.5585E+00   0.1358E+02
SO3--               0.4797E-01   0.3840E+01
SO4--               0.3683E+00   0.3538E+02
CL-                 0.6373E+01   0.2259E+03

TOTAL FLOW RATE = 0.7119E+05 LB/HR
                =       90.   GPM


SUPERNATE RETURN TO SCRUBBER OR EHT
--------- ------ -- -------- -- ---

SPECIES             LB-MOLE/HR   LB/HR

H2O                 0.1076E+05   0.1939E+06
CA++                0.2103E+02   0.8430E+03
MG++                0.3658E+01   0.9378E+02
SO3--               0.3313E+00   0.2652E+02
SO4--               0.2544E+01   0.2443E+03
CL-                 0.4401E+02   0.1560E+04

TOTAL FLOW RATE = 0.1967E+06 LB/HR
                =      393.   GPM


RECYCLE SLURRY TO SCRUBBER
------- ------ -- --------

SPECIES             LB-MOLE/HR   LB/HR

CASO3 .1/2 H2O      0.2455E+05   0.3169E+07
CASO4 .2H2O         0.6037E+04   0.1039E+07
CACO3               0.8449E+04   0.8457E+06
INSOLUBLES          ----------   0.2091E+06
H2O                 0.1632E+07   0.2940E+08
CA++                0.3189E+04   0.1278E+06
MG++                0.5850E+03   0.1422E+05
SO3--               0.5024E+02   0.4022E+04
SO4--               0.3857E+03   0.3705E+05
CL-                 0.6674E+04   0.2366E+06

TOTAL FLOW RATE = 0.3506E+08 LB/HR
                =  63722.     GPM
```

```
FLUE GAS COOLING SLURRY
---- --- ------- ------

SPECIES             LB-MOLE/HR    LB/HR

CASO3 .1/2 H2O      0.1964E+04    0.2535E+06
CASO4 .2H2O         0.4830E+03    0.8312E+05
CACO3               0.6759E+03    0.6766E+05
INSOLUBLES          ----------    0.1673E+05
H2O                 0.1306E+06    0.2352E+07
CA++                0.2551E+03    0.1023E+05
MG++                0.4680E+02    0.1138E+04
SO3--               0.4019E+01    0.3217E+03
SO4--               0.3086E+02    0.2964E+04
CL-                 0.5339E+03    0.1893E+05

 TOTAL FLOW RATE = 0.2807E+07 LB/HR
                 =  [illegible]   GPM
```

```
                    RESULT OF OPTIMIZING POND SIZE TO MINIMIZE TOTAL POND COST AND OVERHEAD

YD3 SLUDGE TO BE DISPOSED OVER LIFE OF PLANT=  7646093. EQUIVALENT DISPOSAL VOLUME IN ACRE FEET.=   4739.
POND DEPTH,FT.= 19.56  EXCAVATION DEPTH.FT.=  3.04
YD3EXC=  1201760. PDACRE=     283. YD2CSD=     93466. YD2ISD=   125850. YD2BOT=  1106403.
PDPERM=    13357. DIVLG=     2436.
CLEAR$=   479490. EXCAV$=  2543580. DIKE$=   905831.
CLAY$=  1026877. HYPLM$=         0. HYPLL$=         0. FERTM$=    44759.
FERTL$=    61892. ROADM$=    16978. ROADL$=      8804. PONDM$=    61737. PONDL$=  5026472.


 POND COST =   5088208.
 LAND COST =    991127.          WITH OVERHEAD =   9539308.
TOTAL COST =   6079335.
```

LIMESTONE SLURRY PROCESS -- BASIS: 500 MW UNIT, 1980 STARTUP

PROJECTED CAPITAL INVESTMENT REQUIREMENTS - EXAMPLE OUTPUT LIMESTONE SCRUBBING WITH ONSITE PONDING

	INVESTMENT, THOUSANDS OF 1979 DOLLARS				DISTRIBUTION
	RAW MATERIAL HANDLING AND PREPARATION	SCRUBBING	WASTE DISPOSAL	TOTAL	PERCENT OF DIRECT INVESTMENT
EQUIPMENT					
MATERIAL	1222.	6195.	51.	7468.	28.2
LABOR	285.	935.	35.	1255.	4.7
PIPING					
MATERIAL	192.	1962.	880.	3034.	11.4
LABOR	90.	637.	365.	1092.	4.1
DUCTWORK					
MATERIAL	0.	1562.	0.	1562.	5.9
LABOR	0.	1187.	0.	1187.	4.5
FOUNDATIONS					
MATERIAL	124.	86.	12.	222.	0.8
LABOR	588.	258.	35.	882.	3.3
POND CONSTRUCTION	0.	0.	5088.	5088.	19.2
STRUCTURAL					
MATERIAL	254.	163.	1.	417.	1.6
LABOR	91.	405.	6.	501.	1.9
ELECTRICAL					
MATERIAL	154.	346.	100.	599.	2.3
LABOR	333.	599.	255.	1187.	4.5
INSTRUMENTATION					
MATERIAL	77.	601.	8.	686.	2.6
LABOR	19.	111.	3.	133.	0.5
BUILDINGS					
MATERIAL	33.	0.	0.	33.	0.1
LABOR	58.	0.	0.	58.	0.2
SERVICES AND MISCELLANEOUS	152.	648.	295.	1095.	4.1
SUBTOTAL DIRECT INVESTMENT	3672.	15694.	7133.	26498.	100.0
ENGINEERING DESIGN AND SUPERVISION	330.	1412.	642.	2385.	9.0
CONSTRUCTION EXPENSES	587.	2511.	1141.	4240.	16.0
CONTRACTOR FEES	184.	785.	357.	1325.	5.0
CONTINGENCY	367.	1569.	713.	2650.	10.0
SUBTOTAL FIXED INVESTMENT	5140.	21971.	9986.	37098.	140.0
ALLOWANCE FOR STARTUP AND MODIFICATIONS	411.	1758.	799.	2968.	11.2
INTEREST DURING CONSTRUCTION	617.	2637.	1198.	4452.	16.8
SUBTOTAL CAPITAL INVESTMENT	6168.	26365.	11984.	44517.	168.0
LAND	7.	2.	1000.	1009.	3.8
WORKING CAPITAL	142.	608.	276.	1026.	3.9
TOTAL CAPITAL INVESTMENT	6318.	26975.	13260.	46552.	175.7

```
LIMESTONE SLURRY PROCESS -- BASIS:  500 MW UNIT, 1980 STARTUP

PROJECTED REVENUE REQUIREMENTS - EXAMPLE OUTPUT LIMESTONE SCRUBBING WITH ONSITE PONDING

                                  DISPLAY SHEET FOR YEAR=    1
                                 ANNUAL OPERATION KW-HR/KW =  7000

                      26.76 TONS PER HOUR                         DRY              SLUDGE
                           TOTAL FIXED INVESTMENT       46552000                            TOTAL
                                                                                           ANNUAL
                                                  ANNUAL_QUANTITY        UNIT_COST,$       COST,$
    DIRECT_COSTS

      RAW_MATERIAL

        LIMESTONE                                    149.5 M TONS        7.00/TON         1046500
        LIME                                           0.0 M TONS       40.00/TON        ________0

          SUBTOTAL RAW MATERIAL                                                          1046500

      CONVERSION_COSTS

        OPERATING LABOR AND
          SUPERVISION                             25990.0 MAN-HR        12.50/MAN-HR       324800
        UTILITIES
          STEAM                                  641620.0 M LB           2.00/M LB        1283200
          PROCESS WATER                          252810.0 M GAL          0.12/M GAL         30300
          ELECTRICITY                          48406260.0 KWH            0.029/KWH        1403800
        MAINTENANCE
          LABOR AND MATERIAL                                                              1865500
        ANALYSES                                   3760.0 HR            17.00/HR        ____63900

          SUBTOTAL CONVERSION COSTS                                                       4971500

          SUBTOTAL DIRECT COSTS                                                           6018000

    INDIRECT_COSTS

      DEPRECIATION                                                                        1483900
      COST OF CAPITAL AND TAXES, 17.20% OF UNDEPRECIATED INVESTMENT                       8007000
      INSURANCE & INTERIM REPLACEMENTS,  2.70% OFTOTAL CAPITAL INVESTMENT                 1256900
      OVERHEAD
        PLANT,  50.0% OF CONVERSION COSTS LESS UTILITIES                                  1127100
        ADMINISTRATIVE, RESEARCH, AND SERVICE,
          10.0% OF OPERATING LABOR AND SUPERVISION                                      ____32500

          SUBTOTAL INDIRECT COSTS                                                       _11907400

          TOTAL ANNUAL REVENUE REQUIREMENT                                               17925400
-------------------------------------------------------------------------------------------------------------------
       HEAT RATE   9000. BTU/KWH     -     HEAT VALUE OF COAL     10500 BTU/LB     -      COAL RATE   1500000 TONS/YR
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LIMESTONE SLURRY PROCESS -- BASIS:  500 MW UNIT, 1980 STARTUP

PROJECTED LIFETIME REVENUE REQUIREMENTS - EXAMPLE OUTPUT LIMESTONE SCRUBBING WITH ONSITE PONDING

                                              TOTAL CAPITAL INVESTMENT:  $   46552000

                                                                                          ADJUSTED GROSS
                                                                                          ANNUAL REVENUE
                                              SULFUR        BYPRODUCT                     REQUIREMENT   TOTAL      NET ANNUAL   CUMULATIVE
                                              REMOVED       RATE,          SLUDGE         EXCLUDING     ANNUAL     INCREASE     NET INCREASE
YEARS ANNUAL   POWER UNIT    POWER UNIT       BY            EQUIVALENT     FIXATION FEE   SLUDGE        SLUDGE     IN TOTAL     IN TOTAL
AFTER OPERA-   HEAT          FUEL             POLLUTION     TONS/YEAR      $/TON          FIXATION      FIXATION   REVENUE      REVENUE
POWER TION,    REQUIREMENT,  CONSUMPTION,     CONTROL                                     COST,         COST,      REQUIREMENT, REQUIREMENT,
UNIT  KW-HR    MILLION BTU   TONS COAL        PROCESS,      DRY            DRY            $/YEAR        $/YEAR     $            $
START /KW      /YEAR         /YEAR            TONS/YEAR     SLUDGE         SLUDGE
----------------------------------------------------------------------------------------------------------------------------------------------
  1   7000     31500000      1500000          35000         187300         0.0            17925400      0          17925400     17925400
  2   7000     31500000      1500000          35000         187300         0.0            17670200      0          17670200     35595600
  3   7000     31500000      1500000          35000         187300         0.0            17415000      0          17415000     53010600
  4   7000     31500000      1500000          35000         187300         0.0            17159700      0          17159700     70170300
  5   7000     31500000      1500000          35000         187300         0.0            16904500      0          16904500     87074800
----------------------------------------------------------------------------------------------------------------------------------------------
  6   7000     31500000      1500000          35000         187300         0.0            16649300      0          16649300    103724100
  7   7000     31500000      1500000          35000         187300         0.0            16394000      0          16394000    120118100
  8   7000     31500000      1500000          35000         187300         0.0            16138800      0          16138800    136256900
  9   7000     31500000      1500000          35000         187300         0.0            15883600      0          15883600    152140500
 10   7000     31500000      1500000          35000         187300         0.0            15628600      0          15628600    167768900
----------------------------------------------------------------------------------------------------------------------------------------------
 11   5000     22500000      1071400          25000         133800         0.0            13678300      0          13678300    181447200
 12   5000     22500000      1071400          25000         133800         0.0            13423100      0          13423100    194870300
 13   5000     22500000      1071400          25000         133800         0.0            13167900      0          13167900    208038200
 14   5000     22500000      1071400          25000         133800         0.0            12912600      0          12912600    220950800
 15   5000     22500000      1071400          25000         133800         0.0            12657600      0          12657600    233608200
----------------------------------------------------------------------------------------------------------------------------------------------
 16   3500     15750000       750000          17500          93700         0.0            11062800      0          11062800    244671000
 17   3500     15750000       750000          17500          93700         0.0            10807500      0          10807500    255478500
 18   3500     15750000       750000          17500          93700         0.0            10552300      0          10552300    266030800
 19   3500     15750000       750000          17500          93700         0.0            10297100      0          10297100    276327900
 20   3500     15750000       750000          17500          93700         0.0            10041800      0          10041800    286369700
----------------------------------------------------------------------------------------------------------------------------------------------
 21   1500      6750000       321400           7500          40100         0.0             7821300      0           7821300    294191000
 22   1500      6750000       321400           7500          40100         0.0             7566100      0           7566100    301757100
 23   1500      6750000       321400           7500          40100         0.0             7310900      0           7310900    309068000
 24   1500      6750000       321400           7500          40100         0.0             7055600      0           7055600    316123600
 25   1500      6750000       321400           7500          40100         0.0             6800400      0           6800400    322924000
----------------------------------------------------------------------------------------------------------------------------------------------
 26   1500      6750000       321400           7500          40100         0.0             6545200      0           6545200    329469200
 27   1500      6750000       321400           7500          40100         0.0             6289900      0           6289900    335759100
 28   1500      6750000       321400           7500          40100         0.0             6034700      0           6034700    341793800
 29   1500      6750000       321400           7500          40100         0.0             5779500      0           5779500    347573300
 30   1500      6750000       321400           7500          40100         0.0             5524200      0           5524200    353097500
----------------------------------------------------------------------------------------------------------------------------------------------

TOT  127500    573750000     27321000         637500        3411500                       353097500     0          353097500
   LIFETIME AVERAGE INCREASE IN UNIT REVENUE REQUIREMENT
                          DOLLARS PER TON OF COAL BURNED                                  12.92         0.0        12.92
                          MILLS PER KILOWATT-HOUR                                          5.54         0.0         5.54
                          CENTS PER MILLION BTU HEAT INPUT                                61.54         0.0        61.54
                          DOLLARS PER TON OF SULFUR REMOVED                              553.88         0.0       553.88
REVENUE REQUIREMENT DISCOUNTED AT  11.6% TO INITIAL YEAR, DOLLARS                     125590700         0        125590700
   LEVELIZED INCREASE IN UNIT REVENUE REQUIREMENT EQUIVALENT TO DISCOUNTED REQUIREMENT OVER LIFE OF POWER UNIT
                          DOLLARS PER TON OF COAL BURNED                                  11.79         0.0        11.79
                          MILLS PER KILOWATT-HOUR                                          5.05         0.0         5.05
                          CENTS PER MILLION BTU HEAT INPUT                                56.16         0.0        56.16
                          DOLLARS PER TON OF SULFUR REMOVED                              505.40         0.0       505.40
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NONREGENERABLE PROCESSES SESSION

Session Chairman

H. William Elder

Manager, Emission Control Development Projects
Tennessee Valley Authority
Muscle Shoals, Alabama

RESULTS OF LIME AND LIMESTONE TESTING WITH FORCED OXIDATION AT THE EPA ALKALI SCRUBBING TEST FACILITY

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Bechtel Corporation
San Francisco, California

ABSTRACT

Forced oxidation of the calcium sulfite reaction product in both lime and limestone FGD systems has been successfully demonstrated in 10 Mw prototype units at the Shawnee Test Facility. The oxidized gypsum product results in less disposal volume and settles by an order-of-magnitude faster than the unoxidized material. It filters to better than 80 percent solids and handles like moist soil compared with the unoxidized material which filters only to about 50 to 60 percent solids and is thixotropic.

Forced oxidation by air sparging in the hold tank of the first two independent scrubbing stages was successfully demonstrated on the venturi/spray tower system. Typical conditions were an air stoichiometry of 1.5 atoms oxygen per mole SO_2 absorbed and a pH of 5.5 in the hold tank. Tests were made both with and without fly ash in the flue gas with equally successful results.

Forced oxidation in a single scrubber loop was successfully demonstrated in the TCA system with limestone slurry. Air/slurry contact in the scrubber loop was achieved by pumping slurry from a small downcomer hold tank through an air eductor to a larger oxidation tank where limestone was added. Slurry from the oxidation tank was returned to the scrubber. Typical conditions were a pH of 5.15 to the eductor, 5.5 in the oxidation tank, and an air stoichiometry of 2.5 atoms oxygen per mole SO_2 removed. Maintaining air/slurry contact in the discharge plume from the eductor proved to be critical for good oxidation.

Forced oxidation of the scrubber bleed slurry was less successful. Residual alkalinity in the bleed slurry solids resulted in a rapid rise in pH to a level where the oxidation rate was too slow to be practical. By using sulfuric acid to maintain the pH below 6.0, batch oxidation of bleed slurry was achieved. However, the enhancement of solids properties was not as great as with forced oxidation within the scrubber loop.

Section 1

INTRODUCTION

In wet lime and limestone scrubbing for sulfur dioxide and particulate removal from boiler flue gas, the waste product is normally a slurry of calcium sulfite, calcium sulfate (gypsum), and fly ash. Studies at the 0.1 Mw EPA pilot plant located at Research Triangle Park, North Carolina have shown that the calcium sulfite can be readily oxidized to gypsum by simple air/slurry contact.(1) The resultant product has improved properties including higher settling rates, improved dewatering characteristics, and reduced total waste volume.

To demonstrate the forced oxidation concept on a larger scale than the EPA-RTP pilot plant, forced oxidation tests are being conducted on the 10 Mw EPA prototype scrubbers located at the TVA Shawnee Power Station, Paducah, Kentucky. This paper presents the results to date of the Shawnee forced oxidation tests.

There are two scrubber systems operating at the EPA sponsored Shawnee Test Facility, each with its own independent slurry handling facilities. Both systems were tested with forced oxidation. The systems have the following scrubbers:

- A venturi followed by a spray tower
  (35,000 acfm capacity @ 300°F)
- A Turbulent Contact Absorber (TCA)
  (30,000 acfm capacity @ 300°F)

The scrubbers receive flue gas from TVA Shawnee coal-fired boiler No. 10. The boiler normally burns a high-to-medium-sulfur bituminous coal producing SO_2 concentrations of 1500 to 4500 ppm. Flue gas can be taken from either side of the Boiler No. 10 particulate removal equipment, allowing testing with high fly ash loadings (4 to 8 grains/scf dry) or low loadings (0.04 to 0.10 grains/scf dry).

The Shawnee Test Facility has been operating since March 1972. Bechtel Corporation of San Francisco is the major contractor and test director; TVA is the constructor and test facility operator. The initial test program lasted through October 1974 (2) with the major emphasis on demonstrating reliable operation. The forced oxidation tests are a part of an advanced test program which is scheduled to continue through February 1978. Earlier results of the advanced test program are reported elsewhere.(3)(4)(5)

The current advanced test program schedule is shown in Figure 1. As can be seen, operation with forced oxidation began on the venturi/spray tower system in January 1977 using two scrubber stages. Operation with forced oxidation on the TCA system began in June 1977 using a single scrubber stage. Bleed stream oxidation tests were conducted from March through May 1977. With a few breaks for special test blocks, operation with forced oxidation should continue through the end of the advanced test program.

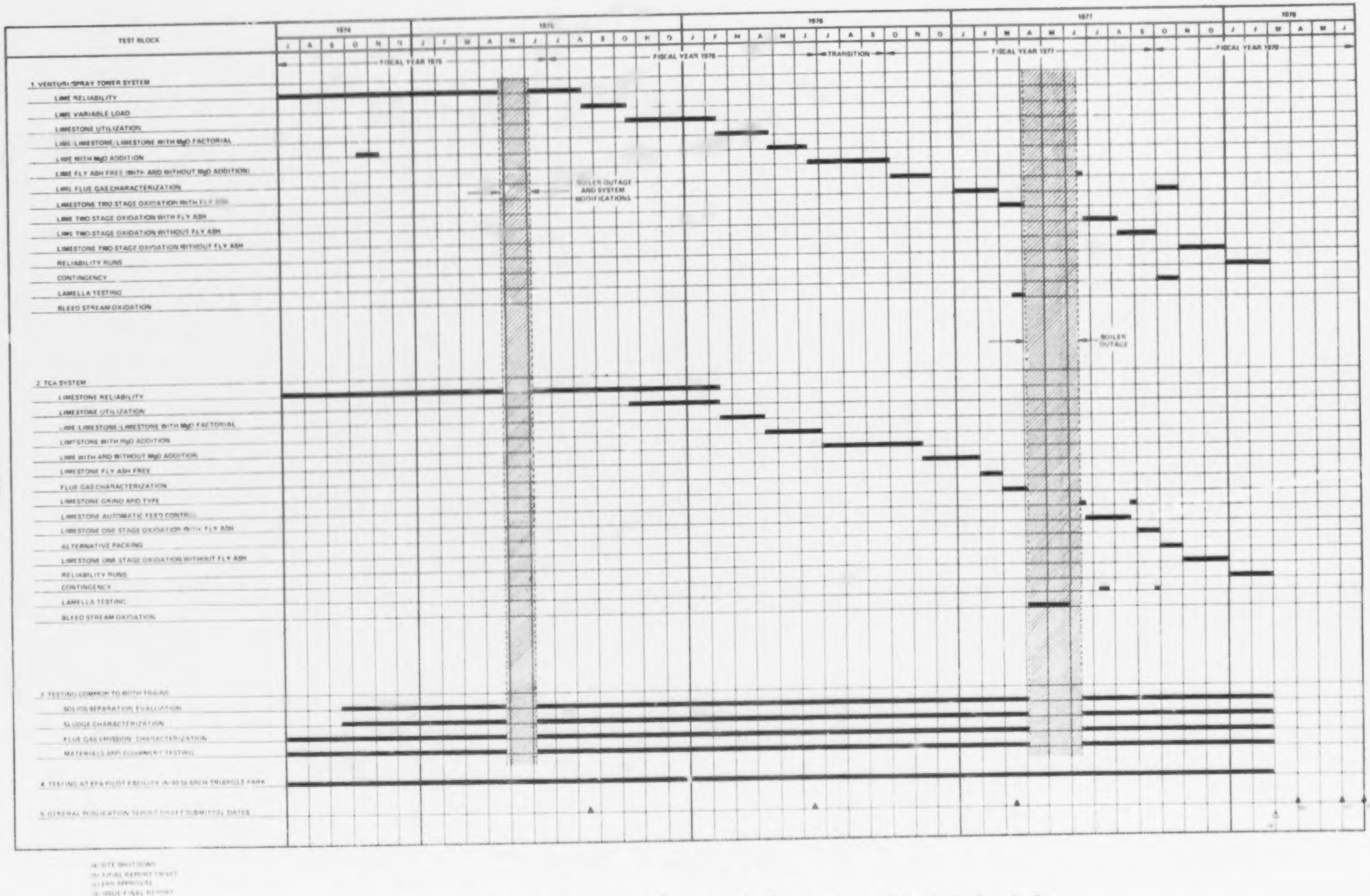

Figure 1. Shawnee Advanced Program Test Schedule.

Section 2

TWO-STAGE FORCED OXIDATION IN THE VENTURI/SPRAY TOWER SYSTEM

Since January 1977, the venturi/spray tower system has operated with two scrubber stages in series and with forced oxidation accomplished by sparging air into the first stage hold tank. Tests were conducted both with lime and limestone slurries and with high and low fly ash loadings in the flue gas. These tests have shown that forced oxidation of sulfite to sulfate (gypsum) is easily achieved in a two-stage scrubber system with an oxidation tank pH range of 4.5 to 5.5.

SYSTEM DESCRIPTION

The venturi/spray tower system was modified for two-stage scrubber operation with forced oxidation as shown in Figure 2. To separate the venturi and spray tower scrubber loops, a catch funnel was installed beneath the bottom spray header of the spray tower. To eliminate slurry entrainment through the catch funnel, the bottom spray header had to be turned upward.

The hold tank in the first scrubber stage (venturi) recirculation loop was used as the oxidation tank. This tank was 8 ft in diameter and operated at either a 14 or 18-ft slurry level. It contained an air sparger ring arranged as shown in Figure 3. The sparger ring was made of straight 3-inch 316L SS pipe pieces welded into an octagon of approximately 4 ft diameter. It was located 6 inches from the bottom of the tank. The initial sparger ring had 130 1/8-inch diameter holes pointed downward. This sparger was later replaced with another one that contained 40 1/4-inch diameter holes. The sparger ring was fed with compressed air to which sufficient water was added to assure humidification. The air must be humidified to prevent local supersaturation and scaling at the sparger holes.

The oxidation tank has an agitator with two axial flow turbines, both pumping downward. Each turbine was 52 inches in diameter and contained 4 blades. The bottom turbine was 10 inches above the air sparger. The agitator rotated at 56 rpm and was rated at 17 brake Hp.

A 10-ft diameter desupersaturation tank, operating at a 5-ft slurry level, followed the oxidation tank to provide time for gypsum precipitation and to provide air-free pump suction.

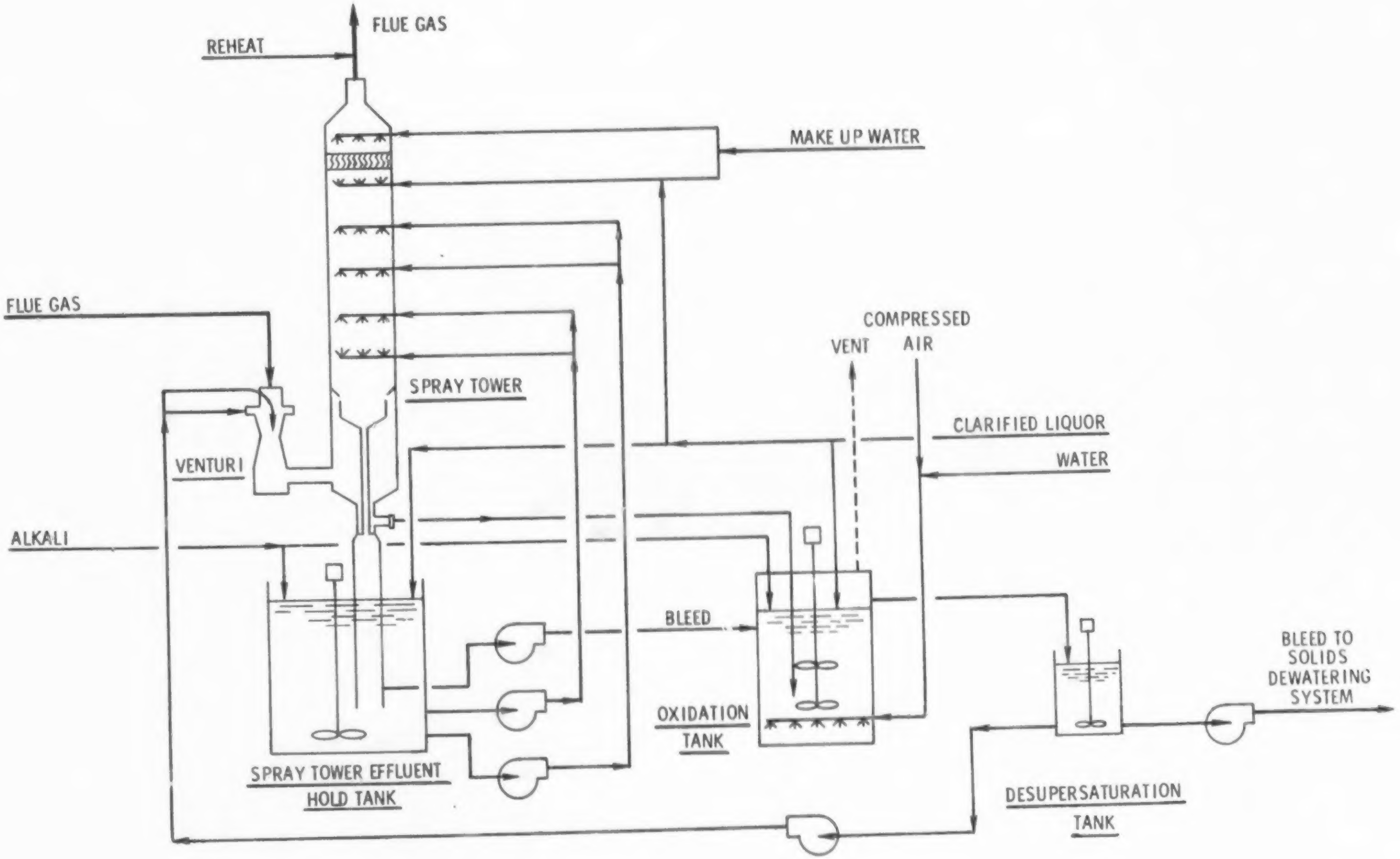

Figure 2. Flow Diagram for Two-Stage Forced Oxidation Tests in the Venturi/Spray Tower System.

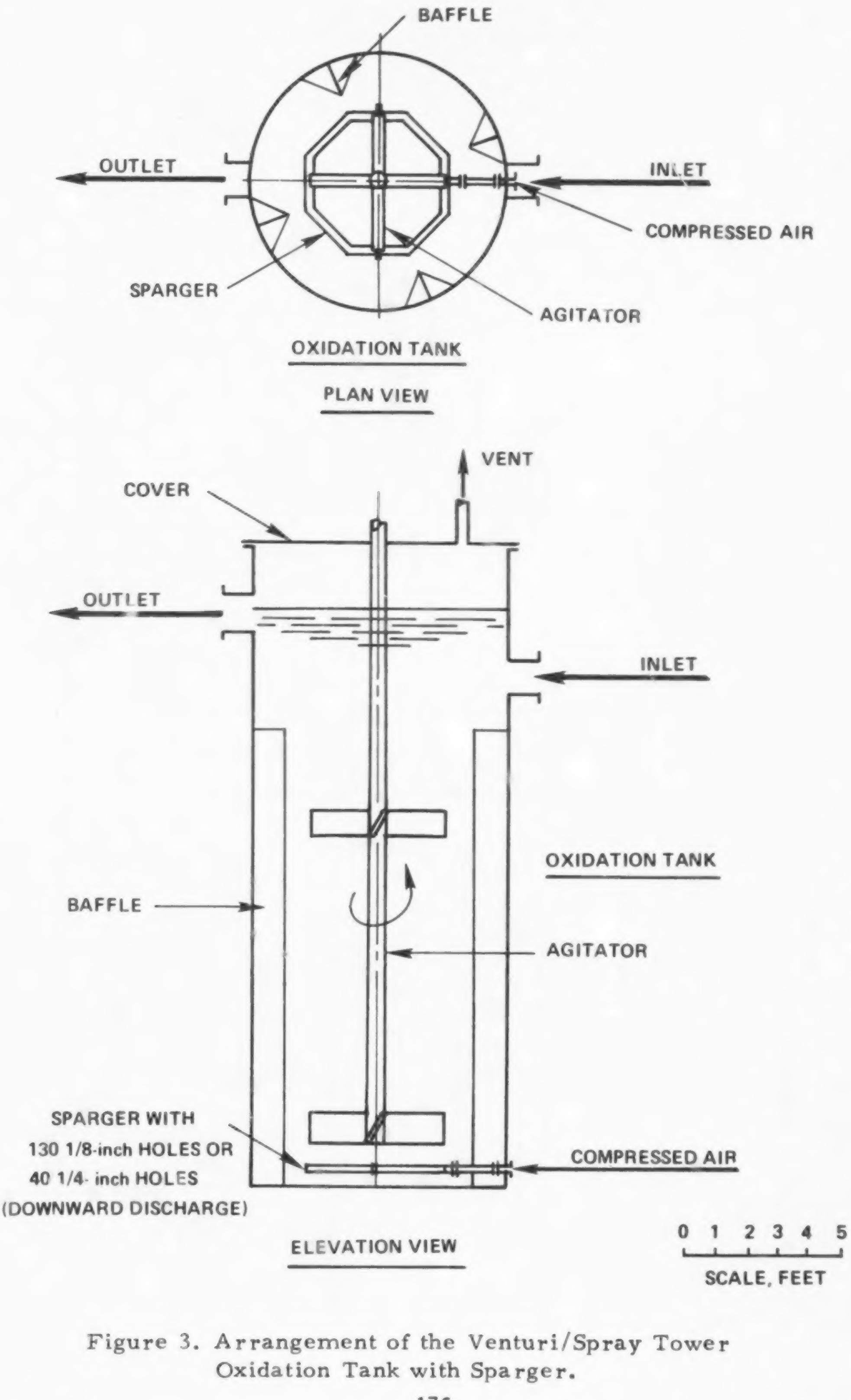

Figure 3. Arrangement of the Venturi/Spray Tower Oxidation Tank with Sparger.

Provision was made to add alkali to either scrubber loop. Clarified liquor from the dewatering system could be returned to either scrubber loop or to the mist eliminator wash circuit.

The first stage (venturi) scrubber loop was fed with slurry bleed from the second stage (spray tower) downcomer. Slurry was bled from the first stage (venturi) scrubber loop to a clarifier followed by a rotary drum vacuum filter for dewatering.

TWO-STAGE FORCED OXIDATION WITH LIMESTONE SLURRY

Forced oxidation tests with two scrubber stages were conducted with limestone slurry from January 4 through March 9, 1977 with high fly ash loadings in the flue gas (Runs 801-1A through 808-1A, Table 1) and again from August 12 through September 15, 1977 with low fly ash loadings (Runs 809-1A through 814-1A, Table 2). Testing with low fly ash loadings is continuing. Tests were conducted at 25,000 acfm gas rate (300°F) which corresponds to a superficial gas velocity in the spray tower of 6.7 ft/sec. A gas rate lower than the maximum used in previous tests (35,000 acfm) was chosen for these runs to assure that high SO_2 removal (greater than 80 percent) could be achieved. Each run averaged about 5 to 6 days which was judged to be sufficient time to reach kinetic equilibrium and to allow adequate run data to be gathered.

Limestone Tests with High Fly Ash Loadings (Table 1)

In the limestone tests with high fly ash loadings, the flue gas was taken from the Boiler 10 duct upstream of the particulate removal equipment. Fly ash loadings in the flue gas ranged from 4 to 8 gr/scf dry. The fly ash was collected in the first scrubber loop (venturi) leaving the slurry in the second loop (spray tower) relatively fly ash-free.

Startup Run. In the initial test with high fly ash loading (Run 801-1A), good sulfite oxidation (93 percent) was achieved by simple air sparging in the venturi stage hold tank. Conditions for Run 801-1A were chosen to approximate those used in tests at the EPA-RTP pilot plant(1) and to provide the best opportunity for oxidation to occur (4.5 pH for optimum reaction rate in the oxidation tank, 18-foot maximum slurry level in the oxidation tank, and an air stoichiometry of about 4 atoms oxygen per mole of SO_2 absorbed). In this run, filter cake solids concentration averaged 81 percent which was the highest ever recorded at Shawnee. Filter cake solids concentrations above 80 percent were consistantly achieved in subsequent runs when sulfite oxidation was above 90 percent. Although oxidation was good in this run, the SO_2 removal was poor (62 percent removal at 3400 ppm inlet).

Table 1

RESULTS OF TWO-STAGE FORCED OXIDATION TESTS ON THE VENTURI/SPRAY TOWER SYSTEM -LIMESTONE SLURRY WITH HIGH FLY ASH LOADINGS-

Major Test Conditions	801-1A	802-1A	803-1A	804-1A	805-1A	806-1A	806-1B	806-1C	806-1D	807-1A	808-1A
Fly ash loading	High	High	High	High	High	High	→	→	→	High	High
Gas rate, acfm @ 300°F	25,000	25,000	25,000	25,000	25,000	25,000	→	→	→	25,000	25,000
Slurry rate to venturi, gpm	400	400	600	600	600	600	→	→	→	600	600
Slurry rate to spray tower, gpm	1300	1400	1400	1400	1400	1400	→	→	→	1400	1400
Venturi percent solids recirculated (controlled)	15	15	15	15	15	15	→	→	→	15	15
Residence times, min: Oxidation tank	17	17	11.3	11.3	11.3	11.3	→	→	→	11.3	11.3
Desupersaturation tank	7	7	4.7	4.7	4.7	4.7	→	→	→	4.7	4.7
Spray tower EHT	19.4	18	18	18	18	18	→	→	→	18	18
Venturi inlet (oxidation tank) pH (controlled)	4.5	5.0	5.0(2)	5.0	5.0	4.5	→	→	→	4.5	4.5
Venturi pressure drop, in. H_2O	9	9	9	9	9	9	→	→	→	9	9
Air rate to sparger, scfm(7)	400	400	400	400	400	250	150	100	50	0	150
Clarified liquor returned to(6)	S. T.	S. T.	S. T.	Vent. & S.T.	Vent.	Vent.	→	→	→	Vent.	Vent.
Selected Results											
Percent SO_2 removal	62	60	67	71	78	73	74	73	80	79	76
Inlet SO_2 concentration, ppm	3400	3500	3150	3150	3450	3400	2850	3250	3100	3350	2850
Spray tower percent solids recirculated	5.8	6.1	5.4	7.4(4)	15.6	15.1	14.5	14.2	15.9	16.6	15.5
Spray tower inlet pH	5.7	5.5	5.65	5.8	6.15	6.15	6.1	6.0	6.15	6.25	6.25
Spray tower limestone stoich. ratio	1.20	1.37	1.47	1.45	1.53	1.30	1.25	1.25	1.35	1.34	1.20
Spray tower inlet liquor gypsum sat'n, %	75	110	120	105	25	20	55	50	15	8	23
Spray tower sulfite oxidation, %	13	16	16	15	11	10	15	17	15	7	16
Overall sulfite oxidation, %	93	95	96	97	98	99	98	97	67	18	97
Overall limestone utilization, %	91	83	87	88	87	95	95	96	91	91	98
Venturi inlet liquor gypsum sat'n, %	100	110	105	105	95	95	95	100	105	135	100
Venturi inlet liquor sulfite concentration, ppm	29	24	21	22	16	16	15	15	130	555	19
Air stoichiometry, atoms O/mole SO_2 abs.	4.7	4.7	4.7	4.4	3.7	2.5	1.7	1.0	0.50	0	1.7
Filter cake solids, wt %(1)	81	86	81	85	81	84	82	79	79(5)	74(5)	82(5)
Mist eliminator restriction, %(8)	2	10	2	1.5(3)	40	-	-	-	50	-	35
Operating hours	263	256	138	151	112	45	15	54	41	66	65

(1) Clarifier and filter in series used for solids dewatering in all runs.
(2) Actual venturi inlet pH averaged 4.8.
(3) Mist eliminator cleaned at 94 on-stream hours (10% restricted).
(4) Controlled by adjusting clarified liquor returned to spray tower EHT.
(5) Value may not be representative due to short duration of run.
(6) Spray tower (effluent hold tank) or venturi (oxidation tank).
(7) Used 130 1/8-inch hole sparger for all runs.
(8) Intermittent mist eliminator bottom wash with makeup water at 1.5 gpm/ft^2 for 4 minutes each hour.

Table 2

RESULTS OF TWO-STAGE FORCED OXIDATION TESTS ON THE VENTURI/SPRAY TOWER SYSTEM -LIMESTONE SLURRY WITH LOW FLY ASH LOADINGS-

Major Test Conditions	809-1A	810-1A	811-1A	812-1A	813-1A	814-1A
Fly ash loading	Low	Low	Low	Low	Low	Low
Gas rate, acfm @ 300°F	25,000	25,000	25,000	25,000	25,000	25,000
Slurry Rate to venturi, gpm	600	600	600	600	600	600
Slurry Rate to spray tower, gpm	1400	1400	1400	1400	1400	1400
Venturi percent solids recirculated (controlled)	15	15	15	15	15	15
Residence times, min: Oxidation tank	11.3	11.3	11.3	11.3	11.3	8.8(5)
Desupersaturation tank	4.7	4.7	4.7	0	0	4.7
Spray tower EHT	13.4	13.4	13.4	13.4	13.4	13.4
Venturi inlet (oxidation tank) pH (controlled)	4.5	5.0	5.5	5.5	5.5	5.5
Venturi pressure drop, in. H_2O	9	9	9	9	9	9
Air rate to sparger, scfm (8)	150	150	150	150	150	150
Clarified liquor returned to (6)	S. T.	S. T.	S. T.	S. T.	S. T.	S. T.
Selected Results						
Percent SO_2 Removal	82	83	85	93		91
Inlet SO_2 concentration, ppm	2450	2700	2600	2350		2450
Spray tower percent solids recirculated	7.9	8.1	8.0	6.9		8.2
Spray tower inlet pH	5.7	5.8	5.9	5.95		5.95
Spray tower limestone stoich. ratio	1.30	1.35	1.40	1.85		1.78
Spray tower inlet liquor gypsum sat'n, %	105	105	105	100		105
Spray tower sulfite oxidation, %	21	22	25	25		26
Overall sulfite oxidation, %	98	98	97	98	(7)	98
Overall limestone utilization, %	98	96	96	80		84
Venturi inlet liquor gypsum sat'n, %	95	95	105	105		105
Venturi inlet liquor sulfite concentration, ppm	35	25	45	67		29
Air stoichiometry, atoms O/mole SO_2 abs.	1.85	1.65	1.65	1.70		1.65
Filter cake solids, wt % (1)	80(3)	-(4)	-(4)	86		88
Mist eliminator restriction, % (2)	0.5	1	2	2.5		4
Operating hours	137	162	184	141	7	136

(1) Clarifier and filter in series used for solids dewatering in all runs except as noted.
(2) Continuous mist eliminator bottom wash with diluted clarified liquor at 0.3 gpm/ft^2.
(3) Excludes last half of run when filter was out of service.
(4) Clarifier only.
(5) Oxidation tank level was 14 ft. All other runs were with 18 ft. level.
(6) Spray tower (effluent hold tank).
(7) Test with oxidation tank agitator turned off. Terminated after 7 hours because the air sparging alone did not keep the solids suspended.
(8) Used 40 1/4-inch hole sparger for all runs.

Improving SO_2 Removal. In Runs 802-1A through 805-1A, operating conditions were changed to improve SO_2 removal. Increasing oxidation tank pH from 4.5 to 5.0 (Run 802-1A) and increasing recirculation rate to the venturi from 400 gpm to 600 gpm (Run 803-1A) improved percent SO_2 removal only slightly (from 62 to 67 percent).

In Runs 804-1A and 805-1A, the solids concentration in the spray tower recirculation loop was increased by returning the clarified liquor from the dewatering system to the venturi loop rather than to the spray tower loop. This change increased the solids in the spray tower loop from 5.4 percent to 15.6 percent with solids controlled at 15 percent in the venturi loop. The higher solids concentration provided more limestone surface to dissolve in the spray tower loop and the spray tower inlet liquor pH increased from 5.65 to 6.15 (pH controlled at 5.0 in the venturi loop). Overall SO_2 removal increased from 67 percent to 78 percent. Subsequent runs were made with clarified liquor return to the venturi loop.

Effect of Air Rate to Oxidizer. Initial runs were made at an air rate to the sparger in the oxidation tank of 400 scfm corresponding to an air stoichiometry of about 4 atoms oxygen per mole SO_2 absorbed. Runs 806-1A through 807-1A were made to determine the minimum air required for near complete sulfite oxidation. The following tests were made at an 18-foot oxidation tank slurry level:

Limestone Slurry

Run	Air Rate, scfm	Air Stoichiometry, atoms O/mole SO_2 abs.	Percent Sulfite Oxidation
806-1A	250	2.5	99
806-1B	150	1.7	98
806-1C	100	1.0	97
806-1D	50	0.5	67
807-1A	0	0	18

The break in oxidation appeared to occur somewhere between 1.0 and 0.5 air stoichiometry. This compares with an air stoichiometry of about 2.6 required in tests at the EPA-RTP pilot plant run under similar conditions without fly ash(1). It has been postulated that an oxidation catalyst, possibly introduced with the fly ash, accounts for the excellent oxidation. Although such a catalyst is suspected, one has not yet been identified in the Shawnee system.

Based on these tests, an air rate of 150 scfm, corresponding to an air stoichiometry of about 1.5 atoms oxygen per mole SO_2 absorbed, was chosen as a standard rate for subsequent runs. Run 808-1A was made at this air rate to confirm the good oxidation.

Limestone Tests with Low Fly Ash Loadings (Table 2)

Following a series of tests with lime slurry, testing with limestone was resumed - this time with low fly ash loadings. The flue gas was taken from the Boiler No. 10 duct downstream of the electrostatic precipitator (Runs 809-1A through 814-1A, Table 2). This flue gas contained particulate concentrations of typically 0.04 to 0.10 gr/scf dry.

Generally, the results with high and low fly ash loadings were similar, with good oxidation and good filter cake being achieved in both cases. During the testing with low fly ash loading, clarified liquor was returned to the spray tower loop for use as part of the mist eliminator wash.

Effect of Oxidation Tank pH. Runs 809-1A through 811-1A were made to determine if the slurry liquor pH in the oxidation tank affected the degree of sulfite oxidation. Over a pH range of 4.5 to 5.5, no adverse effect was seen. Sulfite oxidation remained above 97 percent. Higher pH levels, where a drop in oxidation might be expected, were not investigated as they would be outside of the practical operating range for a two-stage system. The increase in pH also resulted in increased SO_2 removal up to 85 percent in Run 811-1A.

Effect of Desupersaturation Tank. Based on the normal slurry recirculation rate in the venturi loop of 600 gpm, the slurry residence time in the oxidation tank was 11.3 minutes. The oxidation tank overflowed into a desupersaturation tank which provided an additional 4.7 minutes residence time for a total of 16 minutes in the venturi recirculation loop. In Run 812-1A, the desupersaturation tank was removed from the recirculation system with no adverse effect on either sulfite oxidation or gypsum saturation in the venturi loop. However, the desupersaturation tank was returned to service for subsequent runs because it provided a convenient surge for both the recirculation pump and the bleed pump.

Effect of Oxidation Tank Agitator. Run 813-1A was made with the oxidation tank agitator turned off to determine the effect on oxidation of agitating the tank with the sparged air only. The run was aborted after 7 hours because air sparging alone could not keep the solids from settling.

Effect of Oxidation Tank Level. Most runs were made with an 18-foot slurry level in the oxidation tank. However, the tank level was dropped to 14 feet in Run 814-1A and sulfite oxidation was still maintained at 98 percent with no adverse effect on other operating conditions.

TWO-STAGE FORCED OXIDATION WITH LIME SLURRY

Forced oxidation tests with two scrubber stages and lime slurry were conducted from March 10 through August 11, 1977 (Runs 851-1A through

860-1A, Table 3). These tests were interrupted from April 1 through June 21 by a scheduled maintenance outage on Boiler No. 10. Four runs were conducted before the boiler outage with high fly ash loadings and nine runs after the boiler outage with low loadings.

Two-Stage Lime Testing with High Fly Ash Loadings (Table 3)

Initial tests with lime were made at high fly ash loadings (4 to 8 gr/scf dry).

pH Control. In the initial run with lime slurry (Run 851-1A) it became apparent that pH control in the venturi loop was different with lime than with limestone. Residual limestone in the bleed from the spray tower loop to the venturi was sufficient to keep the pH from dropping rapidly in the venturi. The venturi pH was maintained by adding excess limestone to the spray tower loop.

With lime, pH could not be maintained in the venturi under normal operating conditions (600 gpm to the venturi, 9 inches H_2O pressure drop) without raising the pH excessively in the spray tower. In Runs 852-1A and 853-1A, unsuccessful attempts were made to control the venturi pH by changing the recirculation slurry rate and venturi pressure drop. Finally, in Run 854-1A a small amount of lime was added to the venturi loop on pH control. This allowed the pH of the two scrubber loops to be easily controlled independently. Sulfite oxidation was 96 percent on Run 854-1A and SO_2 removal was 82 percent at 3150 ppm inlet SO_2.

Two-Stage Lime Testing with Low Fly Ash Loadings (Table 3)

All additional tests with lime were made with low fly ash loadings (0.04 to 0.10 gr/scf dry).

Effect of Oxidation Tank pH. As with limestone, runs were made (Runs 855-1A through 857-1A) to determine if oxidation tank pH affected sulfite oxidation. Over the pH range of 4.5 to 5.5, no adverse effect on oxidation was observed. During these tests, percent SO_2 removal ranged from 83 to 88 percent at about 2500 ppm inlet. Subsequent runs were made at a venturi inlet pH of 5.5.

Effect of Slurry Solids Concentration. In Run 858-1A, the effect of dropping the venturi loop slurry solids concentration from 15 to 8 percent was investigated. Sulfite oxidation was not affected. It remained high at 97 percent.

In this run, clarified liquor from the dewatering system and makeup water in excess of the mist eliminator wash were added to the venturi

Table 3

RESULTS OF TWO-STAGE FORCED OXIDATION TESTS ON THE VENTURI/SPRAY TOWER SYSTEM -LIME SLURRY WITH HIGH AND LOW FLY ASH LOADINGS-

Major Test Conditions	851-1A	852-1A	853-1A	854-1A	855-1A	856-1A	857-1A	858-1A	859-1A	859-1B	859-1C	859-1D	860-1A
Fly ash loading	High	High	High	High	Low	Low	Low	Low	Low	→	→	→	Low
Gas rate, acfm @ 300°F	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000	→	→	→	25,000
Slurry rate to venturi, gpm	600	160	600	600	600	600	600	600	600	→	→	→	600
Slurry rate to spray tower, gpm	1400	1400	1400	1400	1400	1400	1400	1400	1400	→	→	→	1400
Venturi percent solids recirculated (controlled)	15	15	15	15	15	15	15	8	15	→	→	→	15
Residence times, min: Oxidation tank(6)	11.3	42	11.3	11.3	11.3	11.3	11.3	11.3	11.3	→	→	→	11.3
Desupersaturation tank	4.7	17.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	→	→	→	4.7
Spray tower EHT	18	18	18	18	18	18	18	18	18	→	→	→	12.6
Venturi inlet (oxid. tank) pH (controlled)	4.5	≧ 4.5(3)	≧4.5(4)	5.2	4.5(8)	5.0	5.5	5.5	5.5	→	→	→	5.5
Spray tower inlet pH (controlled)	8.0	8.0	8.0	8.0	8.0	8.0	8.0	8.0	8.0	→	→	→	8.0
Venturi pressure drop, in. H_2O	9	(min)	(min)	9	9	9	9	9	9	→	→	→	9
Air rate to sparger, scfm(2)	150	150	150	150	150	150	150	150	135	150	100	0	150
Clarified liquor returned to(10)	Vent & S. T.	Vent & S. T.	S. T.	S. T.	S. T.	S. T.	S. T.	Vent	S. T.	→	→	→	S. T.
Selected Results													
Percent SO_2 removal	78	70	77	82	83	88	84	83	92	92	92	93	95
Inlet SO_2 concentration, ppm	3300	3400	3450	3150	2350	2500	2650	2750	2950	2700	2700	2600	2200
Spray tower percent solids recirculated	6.0	8.3	6.4	6.1	7.3	7.9	7.9	15.2	7.2	7.6	7.3	7.1	7.0
Spray tower lime stoich. ratio	1.13	1.13	1.14	1.15	1.14	1.16	1.14	1.13	1.16	1.16	1.15	1.17	1.18
Spray tower inlet liquor gypsum sat'n, %	80	50	80	55	95	90	90	85	60	90	90	50	70
Spray tower sulfite oxidation, %	15	13	14	11	18	12	11	14	10	12	20	10	10
Overall sulfite oxidation, %	97	83	97	96	98	97(9)	97	97	76	99	69	30	99
Overall lime utilization, %	98	96	95	97	98	98	99	98	97	99	96	90	99
Venturi inlet liquor gypsum sat'n, %	95	105	100	100	105	95	100	110	95	105	105	115	100
Venturi inlet liquor sulfite concentration, ppm	39	22	35	35	24	36(9)	64	20	99	33	82	85	20
Air stoichiometry, atoms O/mole SO_2 abs.	1.45	1.55	1.40	1.40	1.90	1.65	1.65	1.60	1.20	1.50	1.0	0	1.75
Filter cake solids, wt%(1)	81	73	78	79	75(7)	83(7)	78	81	74	78	71	55	82
Mist eliminator restriction, %(5)	1	-	1	1	-	0.5	0.5	0.5	-	-	-	-	0.5
Operating hours	110	74	161	166	158	209	128	162	115	70	120	72	133

(1) Clarifier and filter in series used for solids dewatering in all runs.

(2) Used 130 1/8-inch hole sparger for Runs 851-1A through 859-1A and 40 1/4-inch hole sparger for Runs 855-1A through 860-1A.

(3) Actual pH range was 4.6-6.2 (avg. 5.6).

(4) Actual pH range was 4.3-5.2 (avg. 4.8).

(5) Intermittent mist eliminator bottom wash with makeup water at 1.5 gpm/ft² for [illegible] minutes every 4 hours.

(6) Agitator flow pattern was changed from radial to axial beginning with Run 855-1A.

(7) Intermittent filter operation.

(8) Actual average pH was 4.7.

(9) Excluded a period 7/2/77-7/5/77 in which oxidation dropped to as low as 62% for unknown reason.

(10) Spray tower (effluent hold tank) or venturi (oxidation tank).

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loop rather than to the spray tower. This resulted in a buildup of the spray tower slurry solids concentration from 8 to 15 percent. With lime slurry, the higher spray tower slurry solids concentration did not result in improved SO_2 removal as had been observed with limestone slurry (see Run 805-1A). This difference between lime and limestone was expected as lime is more soluble than limestone.

Subsequent runs were made at 15 percent venturi slurry solids and with clarified liquor return to the spray tower effluent hold tank.

Effect of Air Rate to Oxidizer. As with limestone, tests were made to determine the minimum air required for near complete sulfite oxidation. Results were as follows:

Lime Slurry

Run	Air Rate, scfm	Air Stoichiometry, atoms O/mole SO_2 abs.	Percent Sulfite Oxidation
859-1B	150	1.5	99
859-1A	135	1.2	76
859-1C	100	1.0	69
859-1D	0	0	30

In these tests, the break from near complete oxidation occurred between 1.2 and 1.5 air stoichiometry, somewhat higher than with limestone but still excellent. A sparger ring with 130 1/8-inch holes was used in the limestone tests. In the lime tests, a ring with 40 1/4-inch holes was used. Both of these spargers plugged, as will be discussed later. Also slurry liquid pH was higher in the lime tests (5.5) than in the limestone tests (4.5). Any combination of these differences may have contributed to the increase in air stoichiometry required for near complete oxidation.

Effect of Spray Tower Hold Tank Residence Time. In the final test with lime slurry (Run 860-1A), the spray tower hold tank residence time was dropped from 18 minutes to 12.6 minutes (from 10' 9" tank level to 7' 6"). The higher tank level had been used previously to keep the bleed pump taking suction from the spray tower downcomer from cavitating. Before this run, the suction point on the downcomer was lowered to allow lower tank level and more reasonable residence time.

This change had no adverse effect on sulfite oxidation or SO_2 removal. Sulfite oxidation was 99 percent and SO_2 removal was 95 percent at 2200 ppm inlet SO_2 concentration.

General Operating Characteristics of the Two-Stage System with Forced Oxidation

Water Balance. Because the dewatering properties of the oxidized sludge are better, less liquor leaves the system with the waste sludge (better than 80 percent solids compared with about 50 to 60 percent solids with unoxidized sludge) and the overall water balance in the slurry system is tighter. The tighter water balance results in higher dissolved solids in the slurry liquor and less available water for mist eliminator wash.

Slurry Solids Control. The slurry solids concentrations in the two scrubber loops are interrelated by the system water balance. The Shawnee tests have normally been run by holding the solids concentration in the venturi loop at 15 percent and letting the spray tower slurry solids concentration float. Spray tower slurry solids concentration has depended primarily on where the clarified liquor from the dewatering system is returned and on the fly ash loading in the flue gas. With 15 percent solids in the venturi loop, typical spray tower slurry concentrations have been as follows:

Clarified Liquor Returned to the	Fly Ash Loading	Percent Solids in Spray Tower Slurry
Venturi loop	High	14-16
Spray tower loop	Low	7-8
Spray tower loop	High	5-6

With limestone, low slurry solids concentration contributes to low SO_2 removal while with lime, it does not. The only adverse effect observed with lime was slight tendency to scale in the bottom of the tower when the solids concentration dropped below 7 percent.

Mist Eliminator Wash. Because of the tight water balance in the scrubber system, limited water was available to wash the three-pass, 316L SS, open-vane chevron mist eliminator. Washing schemes and results for the 3 major test blocks were as follows:

Alkali	Bottom Wash	Top Wash	Spray Tower Alkali Utilization, Moles SO2 Absorbed per mole Ca feed	Mist Eliminator Condition
LS	HI	HI	.6 to .8	fouled
LS	Cont.	LI	.5 to .7	clean
Lime	LI	LI	>.85	clean

Bottom wash rate:

HI - high intermittent - 1.5 gpm/sq. ft. with makeup water for 4 minutes each hour.

LI - low intermittent - 1.5 gpm/sq. ft. with makeup water for 6 minutes every 4 hours.

Cont. - continuous - 0.3 gpm/sq. ft. with clarified liquor and makeup water at 65/35 ratio.

Top wash rate:

HI - high intermittent - makeup water on a 1-hour sequential cycle with one of the six nozzles activated at 0.5 gpm/sq. ft. for 3 minutes every 10 minutes.

LI - low intermittent - makeup water on an 8-hour sequential cycle with one of the six nozzles activated at 0.5 gpm/sq. ft. for 4 minutes every 80 minutes.

The high intermittent bottom wash did not keep the mist eliminator clean with limestone at low alkali utilization in the spray tower loop, but a continuous wash did. With lime at greater than 85 percent utilization, a low intermittent bottom wash was adequate.

Results of these tests and previous tests have shown that the mist eliminator becomes more difficult to keep clean as gas velocity and percent solids in the slurry increase and as alkali utilization decreases.

Sparger Performance - The original sparger, containing 130 1/8-inch diameter holes, was in service from January 4 through April 1, 1977, during which period 1747 operating hours were logged. When the unit shut down in April for a scheduled 8-week boiler maintenance outage, the sparger was removed for inspection. Half of the holes (those on the far side from the air inlet to the sparger) were found to be plugged with scale while most of the remainder were severely eroded. The sparger was replaced with a similar one having 40 1/4-inch diameter holes. This second sparger was put into service on June 21 following the boiler outage.

On September 9 after 1796 operating hours the second sparger was inspected and found to be almost completely plugged. A 3-inch diameter by 6-inch long spool piece connecting the air line to the sparger was found to be severely corroded with holes in several spots. Most of the sparging air must have been emitted through the holes in this spool piece. The spool piece was definitely not 316L SS as there was no corrosion of connecting pipe.

Because of the aforementioned conditions, air was emitted from the sparger mainly on one side of the oxidation tank for a good portion of the testing. Dispersion was accomplished primarily by the agitator. Despite this maldistribution of air, better than 95 percent sulfite oxidation was achieved in all runs with an air rate of at least 150 scfm (approximately 1.5 atoms oxygen per mole SO_2 removal).

Forced Oxidation Excursion - On July 2, 1977 during Run 856-1A, the sulfite oxidation began a steady decline for no apparent reason from about 95 percent down to a low of 62 percent the next day. It then steadily increased back to 95 percent by midnight July 5. During the same period, a similar decline in oxidation was observed on the TCA system during Run 802-2A. On the TCA system, oxidation dropped to 14 percent and then increased back to the run average of 56 percent. No known changes were made in operating conditions during this period. This was the only excursion in oxidation observed during the entire forced oxidation testing program.

Although the cause of the excursion has not yet been determined, it is postulated that either an oxidation inhibitor must have contaminated the system or a decline in the concentration of trace oxidation catalysts must have occurred. Sources of trace elements would be either the coal burned in Boiler No. 10 or the Ohio River water used for makeup. A program of routinely analyzing the scrubber slurry for trace metals, nitrogen, and organic carbon has been initiated to provide background information in case of another excursion.

Chloride Levels - Chlorides enter the scrubber system as HCl in the flue gas. The chlorides are absorbed and concentrated in the scrubber liquor as the major component of the total dissolved solids. Chloride-ion in the scrubber liquor tends to reduce the liquor pH (at constant limestone stoichiometric ratio) and consequently reduce the SO_2 removal efficiency (see Section 5.3.2, Reference 5).

Because of tighter water balance with forced oxidation, chloride-ion concentration has tended to be higher than in previous operation without forced oxidation. Chloride-ion concentration in the venturi loop during forced oxidation testing ranged from 1000 to 7000 ppm, varying mainly with the chloride concentration of the coal burned. Chloride-ion concentrations in the spray tower loop depended on where the clarified liquor from the dewatering system was returned. With clarified liquor returned to the spray tower loop, spray tower chloride-ion concentration averaged about one-half the concentration in the venturi loop. With clarified liquor returned to the venturi loop, the spray tower chloride-ion concentration was about one-fifth the venturi concentration. With the low spray tower chloride-ion concentrations in the latter case, higher pH, and consequently higher SO_2 removal efficiency, was achieved in the spray tower loop.

Section 3

SINGLE-STAGE FORCED OXIDATION IN THE TCA SYSTEM

Beginning in June 1977, the TCA system has operated as a single scrubber stage with forced oxidation. As an alternative to the sparger, an air eductor was used to accomplish air/slurry contact. Tests were conducted with limestone slurry and with high fly ash loadings. Tests with low fly ash loadings will be conducted later. No single-stage forced oxidation tests were made with lime slurry because sulfite is the major SO_2 scrubbing species on a lime scrubber. Forced oxidation in a single stage reduces the sulfite concentration in the slurry liquor, resulting in poor SO_2 removal. Oxidation of sulfite to sulfate (gypsum) with good SO_2 removal was successfully accomplished on the one-stage system using limestone slurry.

SYSTEM DESCRIPTION

An air eductor (Penberthy ELL-10 Special) was used to provide air/slurry contact. The eductor operates by using the energy of the slurry, pumped through the eductor, to aspirate air into an intimate air/slurry mixture. The mixture is discharged into the oxidation tank where the bulk of the oxidation reaction takes place.

This device was capable of educting 600 scfm of air from the atmosphere at zero back pressure with a slurry flow rate of 1600 gpm. The eductor nozzle was made of stellite and the body was made of neoprene lined carbon steel.

Two operating configurations were used on the single-stage tests. In the one-tank configuration (Figure 4) slurry was recirculated from the TCA effluent hold tank (oxidation tank) through the eductor and back into the hold tank. The hold tank was 20 ft in diameter and operated at slurry levels from 6 to 12 feet. Limestone makeup was added either to the hold tank or to the pump suction of the scrubber recirculation line. In the two-tank configuration (Figure 5) a small downcomer tank (7-ft. diameter) was used to receive the slurry from the scrubber. Slurry was then pumped from the downcomer tank through the eductor to the large hold tank where limestone was added. A slurry tie line connecting the two tanks equalized the slurry level in both tanks. The slurry flow through the eductor was maintained higher than the slurry rate to the scrubber. Thus, there was always a slurry backflow through the tie line from the hold tank to the downcomer tank. The two-tank configuration had the advantage that the pH of the slurry passing through the eductor was lower and thus the chemical reaction rate was higher.

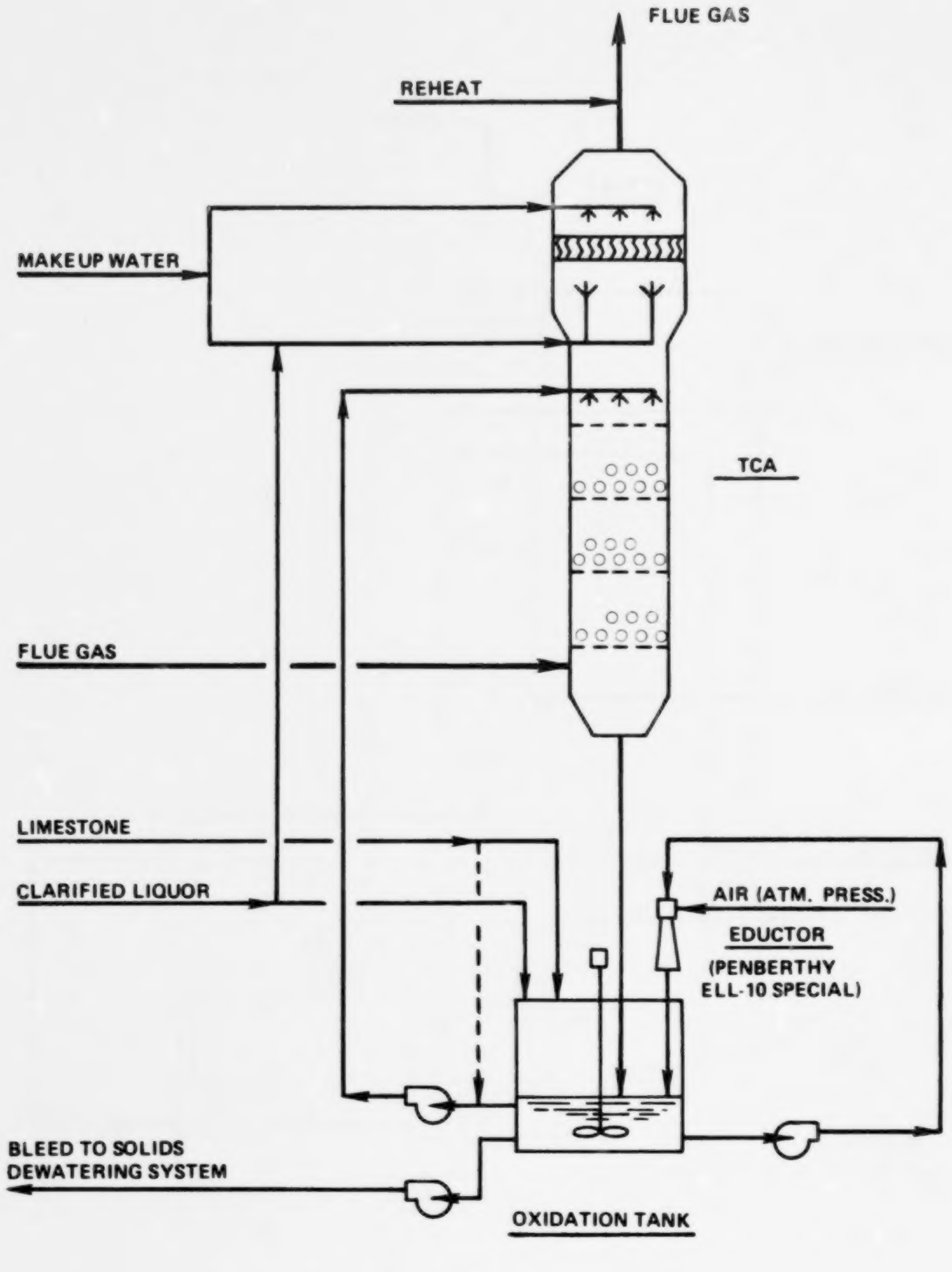

Figure 4. Flow Diagram for Single-Stage Forced Oxidation Tests in the TCA System with One Tank.

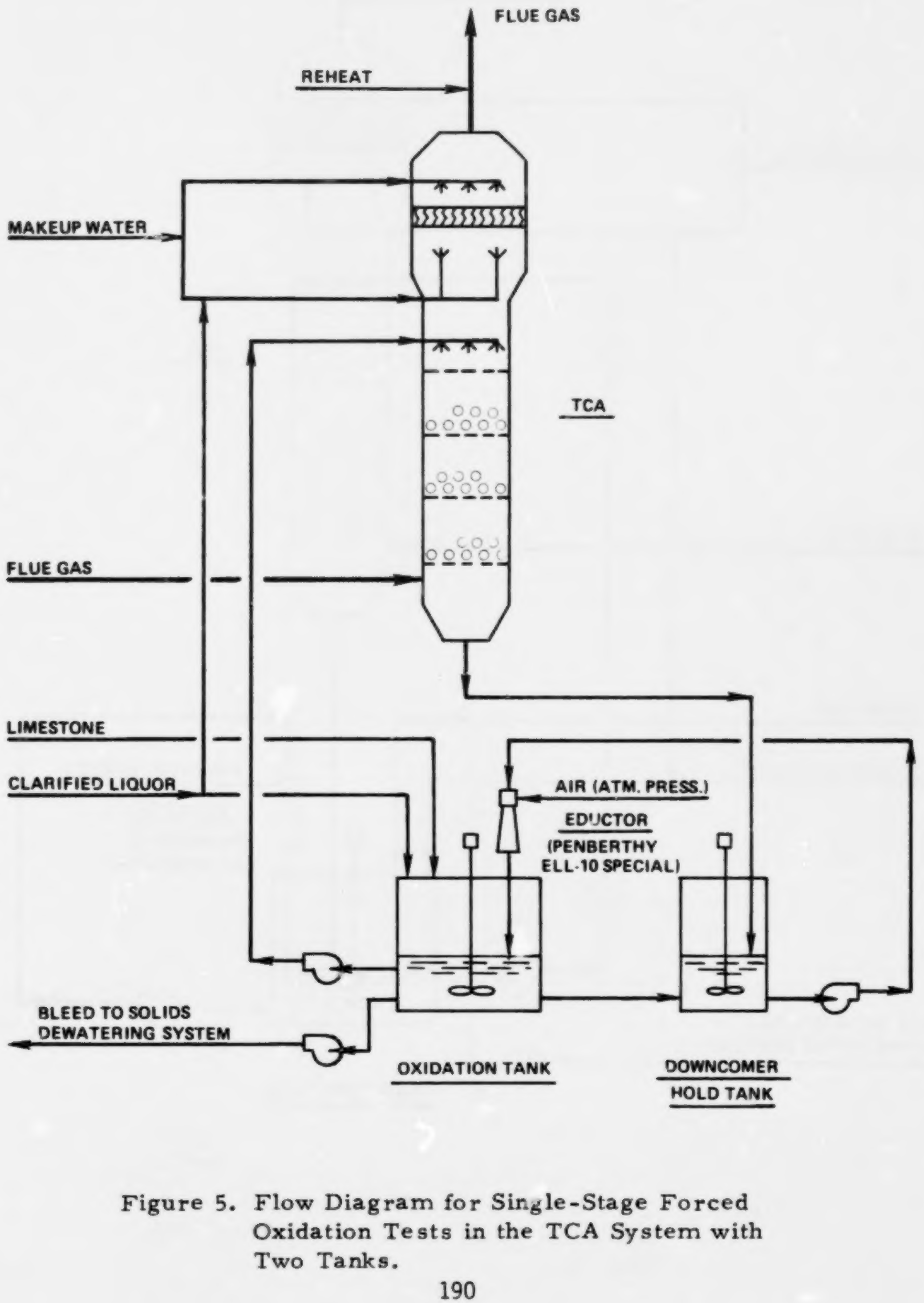

Figure 5. Flow Diagram for Single-Stage Forced Oxidation Tests in the TCA System with Two Tanks.

In both configurations the eductor was oriented to discharged vertically downward into the oxidation tank (see Figure 6).

Normally, slurry was bled from the oxidation tank to a clarifier for dewatering. The clarifier operated with an underflow solids concentration of about 35 to 45 percent. In two runs (809-2A and 810-2A), the clarifier was followed by a rotary drum vacuum filter.

SINGLE-STAGE FORCED OXIDATION WITH LIMESTONE SLURRY

Single-stage forced oxidation tests with limestone slurry and high fly ash loadings in the flue gas (4 to 8 gr/scf dry) were initiated on June 24, 1977 and were continuing as of August 30, 1977. Test results are presented in Table 4. A flue gas rate of 30,000 acfm and a slurry recirculation rate of 1200 gpm were chosen as typical of previous operations without forced oxidation. Each run lasted approximately 5 to 6 days.

Testing in the One-Tank Configuration (Table 4)

Seven runs were made with the TCA in the one-tank, forced-oxidation configuration of Figure 4.

Initial Runs. With a 12 minute residence time in the effluent hold tank, typical of past operation without forced oxidation, the slurry level in the 20 ft diameter tank was only 6.1 feet. This was an awkward height to diameter ratio for forced oxidation. It was difficult to maintain good air/slurry contact long enough to complete oxidation as indicated by the poor oxidtion in the initial runs in the TCA single-stage system.

The first run (Run 801-2A) began with a 6.1 ft slurry level in the hold tank and with the eductor discharging at the slurry surface. With this arrangement, only 53 percent sulfite oxidation was achieved. Oxidation was increased to 63 percent by raising the tank level to 8 feet (Run 801-2B) and discharging the eductor 2 feet below the slurry surface. The 2-foot back pressure on the eductor discharge reduced the educted air rate from 600 scfm to 530.

With the tank level raised to 12 feet and the eductor still discharging at 2 feet below the slurry surface (a 4-ft long spool piece on the eductor discharge was removed), sulfite oxidation was still poor at 56 percent (Run 802-2A).

Both of these runs were made with a slurry liquor pH of about 5.8.

Effect of Low pH. In the next runs (Runs 803-1A and 804-2A) better

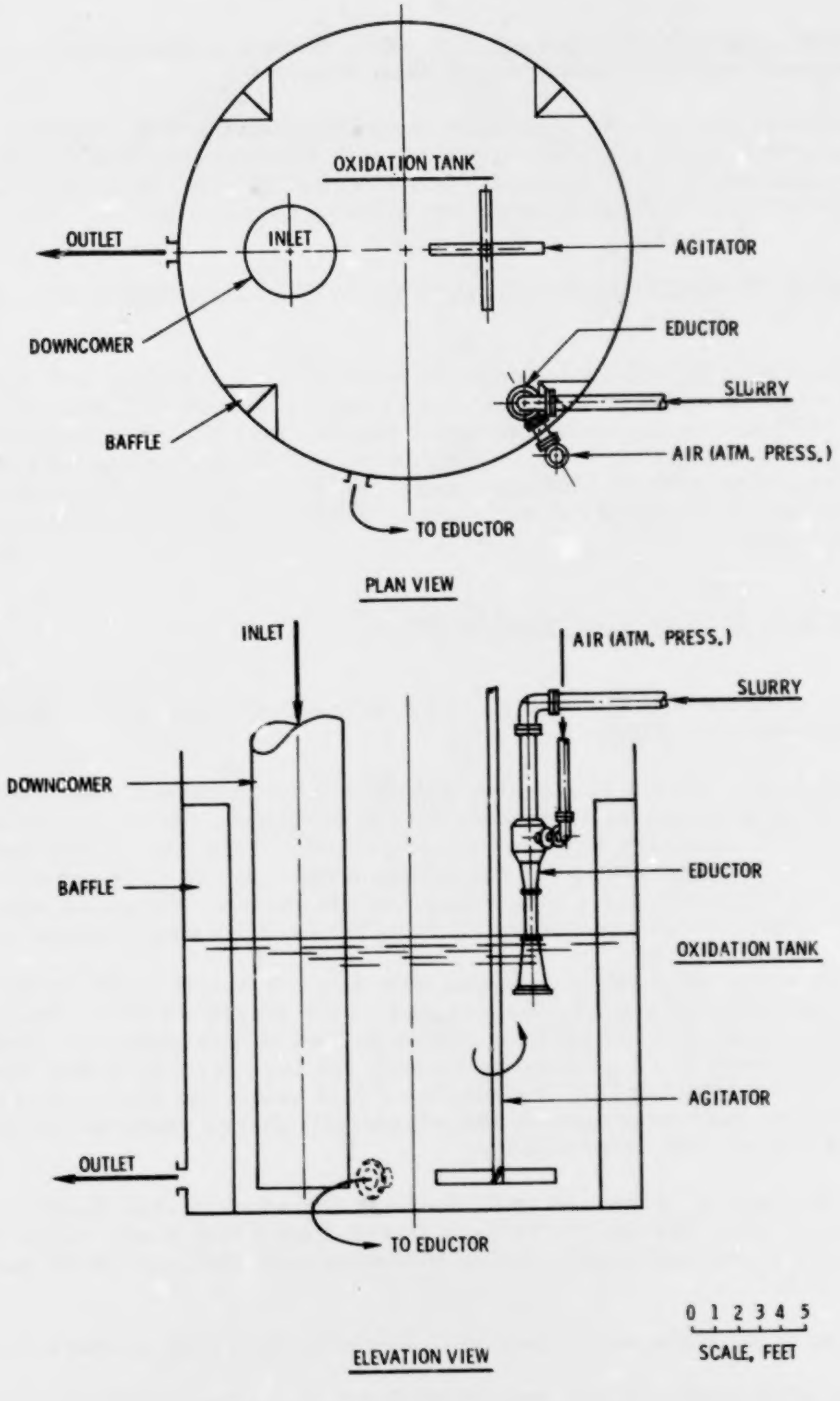

Figure 6. Arrangement of the TCA Oxidation Tank with Eductor.

Table 4

RESULTS OF ONE-STAGE FORCED OXIDATION TESTS ON THE TCA SYSTEM -LIMESTONE SLURRY WITH HIGH FLY ASH LOADINGS-

Major Test Conditions	801-2A	801-2B	802-2A	803-2A	804-2A	805-2A	806-2A	807-2A	808-2A	809-2A(4)	810-2A(4)
Fly ash loading	High	High	High	High	High	High	High	High	High	High	High
Gas rate, acfm @ 300°F	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000	30,000
Slurry flow rate to TCA, gpm	1200	1200	1200	1200	1200	1200	1200	1200	1200	1200	1200
Slurry flow rate to eductor, gpm	1600	1600	1600	1600	1600	1600	1200	1600	1600	1600	1600
Percent solids recirculated	15	15	15	15	15	15	15	15	15	15	15
EHT residence time, min.	12	15.7	23.5	23.5	23.5	23.5	23.5	23.5	23.5	15.7	15.7
Downcomer tank residence time, min.	-	-	-	-	-	-	-	2.9	2.9	1.9	1.9
EHT level, ft	6.1	8.0	12	12	12	12	12	12	12	8	8
Eductor mounting position in EHT	Ver	Ver.	Ver.	Ver.	Ver.	Ver.	Ver.	Ver.	Ver.	Ver.	Ver.
Eductor dischg. point, ft from EHT bottom	6(2)	6(2)	10	10	10	10	10	10	10	6(2)	6(2)
Limestone stoich. ratio controlled at	1.2	1.2	1.2	-	-	-	-	-	1.3	1.3	-
TCA inlet pH controlled at	-	-	-	5.0	5.0	5.3	5.3	5.4	-	-	5.4
EHT agitator speed, rpm	45	45	45	45	68	68	68	68	68	68	68
Limestone addition point(5)	EHT	EHT	EHT	EHT	EHT	TCA	TCA	EHT	EHT	EHT	EH
Air flow rate to eductor, scfm	~600	530	485	465	410	470	290	475	200(3)	300(3)	300(3)
Selected Results											
Percent SO_2 removal	70	70	76	62	62	72	70	83	71	77	84
Inlet SO_2 concentration, ppm	2750	2750	2900	2950	3000	3050	3000	2700	2900	3200	2900
Percent sulfite oxidation	53	63	56	91	93	93	58	96	93	96	98
Air stoich., atoms O/mole SO_2 abs.	6.4	5.6	4.5	5.2	4.5	4.4	2.8	4.3	2.0	2.5	2.5
TCA inlet pH	5.8	5.8	5.75	5.05	5.0	5.35	5.35	5.45	5.1	5.2	5.5
Eductor inlet pH	5.8	5.8	5.75	5.05	5.0	5.25	5.3	5.1	4.95	5.0	5.15
Limestone utilization, percent	80	80	82	90	93	84	91	82	63	63	77
Gypsum sat'n in TCA inlet liquor, percent	105	105	100	105	105	105	110	100	125	115	110
Mist eliminator restriction, percent(1)	-	-	0	-	0	0	0	0	0	0	0
Operating hours	102	47	142	165	166	140	115	185	162	149	130

(1) All runs made with continuous mist eliminator bottom wash with diluted clarified liquor at 0.4 gpm/ft^2.

(2) At discharge of 4 ft by 10-inch diameter pipe attached beneath the eductor.

(3) Air flow controlled by a sliding plate in the air intake line to eductor.

(4) Filter in series with clarifier was used during Runs 809-2A and 810-2A. Filter cake solids contents were 85 and 88%, respectively.

(5) EHT (effluent hold tank), TCA (pump suction on the TCA slurry inlet line).

than 90 percent sulfite oxidation was achieved by dropping the slurry pH to 5.0 and accepting a low SO_2 removal of 62 percent at 3000 ppm inlet. These were the first successful single-stage forced oxidation runs on the TCA system at Shawnee.

Assuming that air/slurry contact time was limited by the poor TCA hold tank configuration, the lower pH must have increased the reaction rate sufficiently for near complete oxidation to take place. At the EPA-RTP pilot plant, where the hold tank had a length to diameter ratio approaching one and the mixing was probably more uniform, no limiting effect of pH on oxidation was observed up to about pH 6.0(1).

Run 803-2A differed from Run 804-2A by an increase in agitator speed from 45 rpm to the maximum obtainable of 68 rpm. The effect was to increase the average oxidation slightly (93 percent) and to reduce time dependent fluctuations.

SO_2 Removal Improvement. To improve SO_2 removal, the limestone addition point was moved from the effluent hold tank to the pump suction on the scrubber inlet feed line and the scrubber inlet pH was increased to 5.3 (Run 805-2A). The net result was an increase in SO_2 removal to 72 percent (at 3050 ppm inlet concentration) with a continuation of 93 percent sulfite oxidation. Addition of the limestone to the scrubber feed line allowed the air/slurry contact to take place at the lowest pH in the scrubber system.

This run was the most successful in the series of limestone runs with a single hold tank. The circulation rate through the eductor was 1600 gpm which gave an air rate of 470 scfm, corresponding to an air stoichiometry of 4.4 atoms oxygen per mole SO_2 absorbed.

Effect of Decreased Slurry Rate to the Eductor. In Run 806-2A, the eductor feed rate was reduced to 1200 gpm with a corresponding reduction in educted air to 290 scfm. Sulfite oxidation dropped to 58 percent, emphasizing again the poor mixing in the oversize effluent hold tank. The air rate was adequate, as indicated by good oxidation in later runs at 1600 gpm eductor feed rate and air rates as low as 200 scfm.

Testing in the Two-Tank Configuration (Table 4)

Four runs were made with the TCA in the two-tank, single-stage, forced-oxidation configuration of Figure 5, all of which achieved better than 90 percent sulfite oxidation.

Effect of TCA Inlet pH. The pH of the TCA slurry inlet (also the oxidation tank pH) was varied from 5.1 to 5.5 while the eductor inlet pH changed only from 4.95 to 5.15. No systematic effect on sulfite oxidation was observed but SO_2 removal increased from 71 percent to 84 percent as the pH was increased.

Effect of Air Rate. Air rate to the eductor was dropped from 475 scfm in Run 807-2A to 200 scfm in Run 808-2A with only a 3 percent reduction in sulfite oxidation (96 percent oxidation reduced to 93 percent). Air rate was controlled at a constant slurry rate of 1600 gpm by restricting the eductor air inlet.

The 200 scfm air rate, corresponding to an air stoichiometry of 2.0 atoms oxygen per mole SO_2 absorbed, was the lowest air rate in the single stage oxidation series in which better than 90 percent oxidation was achieved.

Effect of Slurry Tank Level. Of the four runs with the two-tank configuration, two were at a 12-ft oxidation tank level (23.5 minute residence time) and two were at an 8-ft tank level (15.7 minute residence time). Sulfite oxidation occurred equally as well at the low tank level (96 and 98 percent) as at the high level (96 and 93 percent). These runs were made over a range of slurry pH to the eductor of 4.95 to 5.15.

GENERAL OPERATING CHARACTERISTICS OF THE SINGLE-STAGE SYSTEM WITH FORCED OXIDATION

Water Balance. On the TCA system, forced oxidation did not significantly change the water balance. Only a clarifier was used for dewatering and the undeflow waste solids concentration was run at 35 to 45 percent, the same as without forced oxidation.

During Runs 809-2A and 810-2A, however, the rotary drum vacuum filter was used in series with the clarifier and the oxidized filter cake averaged over 85 percent solids concentration. During these runs, the water balance was tighter than normal without forced oxidation.

Mist Eliminator Wash. During the entire forced oxidation testing period of 1503 on stream hours, there was absolutely no evidence of solids restriction in the three-pass, 316L SS, open-vane chevron mist eliminator. The mist eliminator was washed continuously on the bottomside at a rate of 0.4 gpm/sq. ft. with all the available makeup water plus sufficient clarified liquor from the dewatering system to achieve the desired flow rate. The ratio of clarified liquor to makeup water averaged about one to one. The topside was washed with makeup water on a 1-hour sequential cycle with one of the six nozzles activated at 0.5 gpm/sq. ft. for 3 minutes every 10 minutes.

During these tests, the superficial gas velocity in the scrubber was 12.5 ft/sec, the slurry solids were maintained at 15 percent, and the limestone utilization ranged from 63 to 93 percent.

Eductor Performance. Air eduction by the eductor was as predicted by the manufacturer. The air eduction rate was sensitive to the back pressure (i.e. slurry submergence of the eductor), ranging from 600 scfm at zero submergence to 200 scfm at 12 feet submergence at a 1600 gpm slurry rate. At 1600 gpm, the slurry pressure drop across the eductor was 38 psi. As seen from the two-tank runs on the TCA, better than 90 percent sulfite oxidation could be achieved at an air rate as low as 200 scfm.

The major operational problem in using the eductor for air/slurry contact resulted from the configuration of the eductor discharge hold tank. Most of the oxidation takes place in the hold tank because the residence time in the eductor is extremely short. The 20-ft. diameter hold tank was too large and normally operated at only 6.1 ft. slurry level to achieve 12 minutes residence time. This slurry level was too shallow to allow time for the oxidation reaction to take place. At 8 and 12 ft slurry levels, bettter than 90 percent oxidation could be achieved, but only by reducing pH to increase the reaction rate.

If the eductor were discharged into a smaller but deeper tank, such as the oxidation tank in the venturi/spray tower system, good oxidation could probably be achieved at higher pH and at lower slurry rate.

The eductor body was constructed of neoprene lined carbon steel. The nozzle was made of stellite. The rubber lined body was not satisfactory in slurry service with high fly ash loadings. After 1500 hours, the rubber had eroded in a circular pattern, presumably at the point of impact of the slurry from the nozzle. In a few spots, the rubber had worn to bare steel. Repair with Epoxylite 203 was unsuccessful. With the rubber gone, the bare steel will probably erode quickly. The stellite nozzle showed only minor evidence of erosion.

Enhancement of SO_2 Removal. SO_2 removal during the single-stage forced oxidation tests has tended to be a few percentage points higher than predicted by the SO_2 removal model fitted to previous Shawnee TCA limestone runs without forced oxidation(5). The reason for this removal enhancement has not yet been determined. However, one speculation is that limestone dissolution rate has been increased because of the reduction of carbonate in the liquor by air stripping of CO_2 and the reduction of bisulfite by oxidation. Evaluation of this phenomenon is continuing.

Chloride Levels - Chloride-ion concentration during the TCA limestone forced oxidation tests ranged from 1000 to 3000 ppm which is typical of previous tests without forced oxidation.

Section 4

BLEED STREAM AND BATCH SLURRY OXIDATION TESTS

An alternative to forced oxidation within the scrubber loop would be to oxidize the bleed stream from the scrubber. Such an arrangement would be especially desirable for retrofitting forced oxidation to existing commercial lime/limestone flue gas desulfurization facilities because it would have the least effect on the scrubber operation. The only interaction would be a change in the composition of the liquor returned from the waste solids dewatering system.

However, tests at the EPA-RTP pilot plant (1) showed that forced oxidation of a scrubber bleed stream required higher air stoichiometry compared with oxidation within the scrubbing loop (5.3 versus 3.5). Furthermore, the properties of the oxidized solids were inferior to those obtained by oxidation within the scrubbing loop.

Additional bleed stream oxidation tests at the Shawnee Test Facility confirmed these observations. Tests were conducted with the system shown in Figure 7. A Penberthy ELL-3 air eductor (120 gpm slurry, 30 to 57 scfm air) was used for air/slurry contact. Bleed rate to the oxidation tank was about 3 ppm. Three tests were conducted with limestone slurry with a high fly ash loading bled from the TCA system and three with lime slurry with a low fly ash loading bled from the venturi/spray tower system.

In all tests, the pH of the slurry recirculating through the eductor rose to a level of 7 to 8 and no oxidation occurred. Evidently, the dissolution rate of solid $CaSO_3$ at such a pH level was too slow for significant sulfite oxidation to occur.

In a series of batch oxidation tests conducted on the same equipment during the boiler outage of April - May 1977, complete oxidation was achieved by continuously adding 93 percent sulfuric acid to maintain a slurry pH of 5 to 6 in the oxidation tank. Typically, the oxidation rate was 2 to 3 x 10^{-4} g-mole sulfite oxidized per liter per minute. However, sulfuric acid addition would not seem to be a commercially desirable procedure.

Only marginal improvement was observed in slurry settling rate, settled density, and filter cake solids content of the oxidized slurry from the batch tests. The marginal improvement is in good agreement with the results observed in the EPA-RTP pilot plant bleed stream tests.

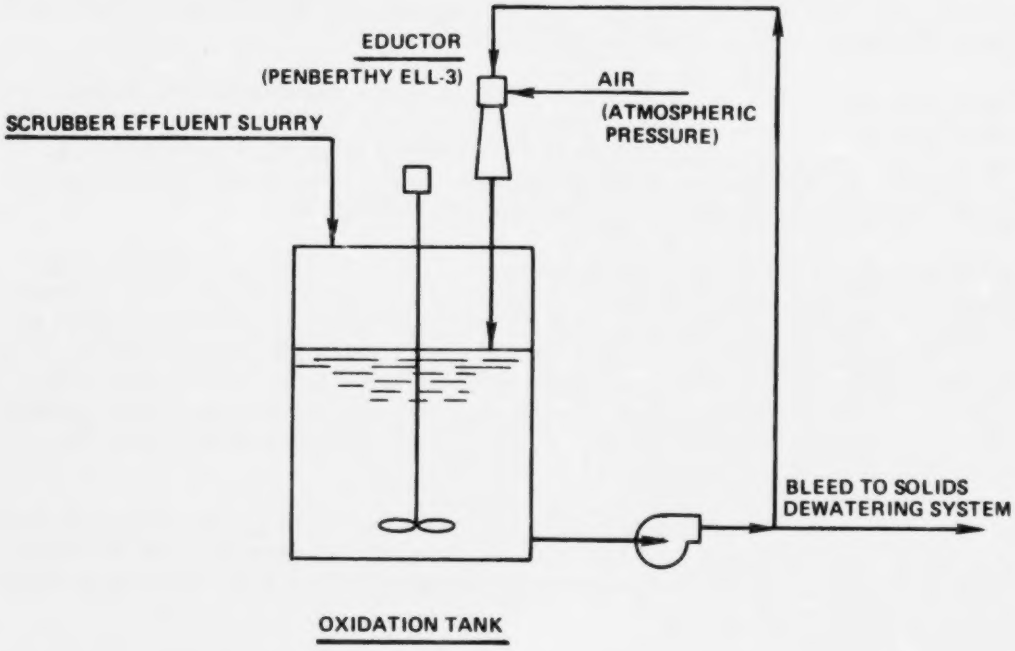

Figure 7. Flow Diagram for Bleed Stream Forced Oxidation Tests.

Section 5

DEWATERING CHARACTERISTICS OF THE OXIDIZED SLURRY SOLIDS

The primary purpose for oxidizing the scrubber solids is to improve waste solids dewatering and disposal characteristics. Dewatering characteristics are routinely monitored at the Shawnee Test Facility by cylinder settling tests and vacuum funnel filtration tests. Cylinder settling and funnel filter tests have been completed and evaluated for all forced oxidation test blocks reported in this paper except for two-stage oxidation with limestone slurry and low fly ash loadings (Runs 809-1A through 814-1A). Results of these tests will be reported later.

Cylinder settling tests are performed in a 1000 ml cylinder containing a rake which rotates at 0.17 rpm. The initial settling rate and ultimate settled solids concentration are recorded as indices of dewatering characteristics. The initial settling rate is a qualitative index of the solids settling properties only. Design rates for sizing clarifiers must take into consideration the hindered settling rate as the solids concentrate. The ultimate settled solids from the cylinder tests represent the highest achievable solids concentration in a settling pond.

Funnel filter tests are performed in a Buchner funnel with a Whatman 2 filter paper under a vacuum of 25 in. Hg. The funnel tests correlate well with the Shawnee rotary drum vacuum filter when not blinded but the funnel test cakes tend to have lower solids concentrations.

Table 5 summarizes the findings to date with respect to the alkali, the fly ash loading, the oxidation method (one or two scrubber stages), and, in the case of the two-stage limestone testing, the pH.

The benefits of forced oxidation on the settling and dewatering characteristics are clearly evident from Table 5. Without forced oxidation, the initial solids settling rate rarely exceeded a few tenths of one cm/min, regardless of alkali type or fly ash content. With forced oxidation, settling rates were generally greater than one cm/min, regardless of the oxidation scheme (one or two scrubber stages).

Similar results were obtained with the ultimate settled solids and the funnel test cake solids. Without forced oxidation, the ultimate settled solids were generally in the range of 40-50 weight percent, while with forced oxidation, the range was 70-80 weight percent solids. Funnel test results indicated 45-55 weight percent solids without forced oxidation and 65-85 weight percent with forced oxidation. On the rotary drum vacuum filter used in the scrubber dewatering system, cake solids concentration was always above 80 percent with oxidized slurry while averaging about 50 to 60 percent with unoxidized slurry. The unoxidized

Table 5

SUMMARY OF THE DEWATERING CHARACTERISTICS OF SHAWNEE SCRUBBER SOLIDS

Alkali	Fly Ash Loading	Forced Oxidation	No. of Scrubber Stages	Initial Settling Rate, cm/min		Ultimate Settled Solids, wt. %		Funnel Test Cake Solids, wt. %		Initial Solids Conc., %
				Avg.	Range	Avg.	Range	Avg.	Range	
LS	High	No	1	0.08	0.04-0.16	48	43-53	--	--	15
LS	Low	No	1	0.10	0.06-0.41	51	46-67	50	43-56	15
LS	High	Yes	1	1.06	0.63-1.27	74	67-84	76	73-90	15
LS	High	Yes	2	1.20[2] 0.70[3]	0.96-1.41[2] 0.55-0.87[3]	73	65-80	69	64-74	15
L	High	No	1	0.20	0.19-0.49	50	48-66	53	51-55	15
L	Low	No	1	0.35	0.09-0.87	40	30-55	45	40-50	8
L	High	Yes	2	1.00	0.70-1.30	78	71-85	68	64-71	15
L	Low	Yes	2	1.45	0.73-2.44	72	64-87	75	64-85	15

(1) Values for forced oxidation runs are only from data where solids oxidation was greater than 90 percent.

(2) Oxidation Tank pH = 5.0.

(3) Oxidation Tank pH = 4.5.

solids tended to be thixotropic, like quicksand, while the oxidized solids were more like moist soil.

For limestone slurry with high fly ash loading, the initial settling rate for solids in slurries with two-stage forced oxidation was significantly lower at an oxidation tank pH of 4.5 than 5.0. At a pH of 5, the average initial settling rate was 1.20 cm/min but at a pH of 4.5 the average rate dropped to 0.70 cm/min. This reduction may have been due to a change in crystal growth patterns at the reduced pH or to a change in the size distribution of the fly ash fines. It is currently considered that the fly ash fines content is the limiting factor in the maximum initial settling rate and not the calcium sulfate fines generated during forced oxidation testing.

Generally, there is a significant reduction in the quality of slurry settling and dewatering characteristics when the oxidation drops below about 90 percent, although the data are sparse for some of the test blocks.

The data thus far generated clearly illustrate the benefits of forced oxidation with respect to increased initial solids settling rate, ultimate settled solids, and funnel test cake solids.

Section 6

CONCLUSIONS

Following the development of forced oxidation techniques at the EPA-RTP pilot plant on a 0.1-Mw scale, forced oxidation has been successfully demonstrated at the Shawnee Test Facility on 10-Mw prototype scrubbers. Based on the Shawnee tests the following conclusions were made:

- Forced oxidation of the slurry within the scrubber loop of both the two-stage venturi/spray tower system and single-stage TCA system dramatically improved the dewatering characteristics of the waste solids.

- Forced oxidation of the slurry batches external to the scrubber recirculation loop required sulfuric acid addition to control pH. Oxidation only marginally improved the solids dewatering characteristics.

- Oxidation of the sulfite solids to gypsum of 90 percent or better was required for maximum improvement of dewatering characteristics.

- Forced oxidation was achieved by simple air/slurry contact at atmospheric pressure in the scrubber hold tank.

- Slurries with high or low fly ash loadings appeared to oxidize equally well.

- Forced oxidation in the first of two independent scrubbing loops was successfully demonstrated in the two-stage venturi/spray tower system with both lime and limestone slurries. Under the base case conditions of 25,000 acfm gas rate, 600 gpm venturi slurry rate, 1400 gpm spray tower slurry rate, 15 percent venturi recirculated slurry solids (with high or low fly ash loadings), and 18 ft oxidation tank level, an average sulfite oxidation better than 96 percent was achieved at an oxidation pH range of 4.5 to 5.5 and an air stoichiometric ratio of about 1.5 atoms oxygen/mole SO_2 absorbed. In these tests an air sparger, discharging directly below a turbine agitator, was used for air/slurry contact.

- Limestone utilization was improved by using two scrubber stages. Under the above conditions, utilization was consistantly better than 90 percent with SO_2 removals up to 91 percent.

- For limestone scrubbing, the venturi inlet pH (oxidation tank pH) was controlled by limestone addition to the spray tower hold tank. For lime scrubbing, however, it was necessary to control both the venturi inlet pH and spray tower inlet pH by separate lime additions to avoid wide pH fluctuations in the venturi and spray tower slurry loops.

- In limestone scrubbing, a low slurry solids concentration reduced the percent SO_2 removal.

- Calcium sulfite scaling in the spray tower was observed when the spray tower recirculated slurry solids concentration dropped below about 7 percent. (The spray tower slurry contained no fly ash in two-stage forced oxidation testing).

- Forced oxidation in the hold tank of the single-stage TCA system was successfully demonstrated with limestone slurry. Under the base conditions of 30,000 acfm gas rate, 1200 gpm slurry rate, 15 percent recirculated slurry solids (with high fly ash loadings) and 8 ft oxidation tank level, an average sulfite oxidation better than 96 percent was achieved at an oxidation tank pH range up to 5.5 and an air stoichiometric ratio of about 2.5 atoms oxygen/mole SO_2 absorbed. In these tests, an air eductor was used for air/slurry contact. Slurry was fed to the eductor at a pH of 5.15 from a small downcomer tank.

- SO_2 removal was enhanced slightly by single-stage forced oxidation with limestone scrubbing.

- An intermittent mist eliminator wash with makeup water was unsatisfactory with limestone slurry at 60 to 80 percent utilization. A continuous wash with diluted clarified liquor was satisfactory under these conditions. The intermittent wash with makeup water was satisfactory with lime slurry at 85 percent utilization.

- Satisfactory oxidation was achieved with an air eductor even though it discharged into a tank having a bad height-to-diameter ratio.

- Long term reliable operation of the sparger or the eductor has not yet been demonstrated.

- Up to 75 percent of the holes in the air sparger plugged with no apparent adverse effect on oxidation.

- The rubber lining of the eductor diffuser eroded severely in less than 1500 hours of operation.

Section 7

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EFFECT OF FORCED OXIDATION ON LIMESTONE/SO_x SCRUBBER PERFORMANCE

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ABSTRACT

Tests conducted at EPA's IERL/RTP pilot plant have indicated the feasibility of oxidizing calcium sulfite slurry to gypsum within the scrubbing loop of single-stage limestone scrubbers at normal operating pH. This paper reports comparative scrubber tests with and without forced oxidation to evaluate its effect on SO_2 absorption efficiency. Also of major concern in these tests was the effect of high chloride concentrations associated with the more tightly closed loop that results from the better dewatering properties of the oxidized sludge. The results of the comparisons show that the principal measures of scrubber performance (i.e., SO_2 removal, limestone utilization, and scrubberfeed supersaturation) will not be adversely affected by the oxidation process: good performance can be expected with 98 percent oxidation and with chloride levels to at least 20,000 ppm. Techniques of aerating the slurry in the scrubber-effluent hold tank are discussed and observations are reported regarding the effect of aeration on the limestone dissolution rate and slurry composition.

EFFECT OF FORCED OXIDATION ON LIMESTONE/SO_x SCRUBBER PERFORMANCE

INTRODUCTION

U.S. electric utilities are currently operating, have under construction, or are planning to install 119 flue gas desulfurization systems representing 50,000 MW of generating capacity. The majority of these systems employ lime or limestone scrubbers to fix SO_2 as calcium sulfite, which can be disposed of in an environmentally acceptable manner. The impending conversion of many oil-fired boilers to coal is likely to further accelerate scrubber construction, and the same considerations forcing this change will probably also favor the use of limestone--rather than the more energy intensive lime--as the fixing agent. The relative appeal of limestone scrubbers is further enhanced by several advances that have improved their performance relative to lime scrubbers; undoubtedly the most important of these advances is the resolution of problems involving mist eliminator fouling(1) which plagued early limestone systems. Another factor is the increasing levels of limestone utilization at which the scrubbers can be operated; limestone utilizations of about 90 percent are now being achieved with both the single-loop TCA and double-loop venturi/spray tower at the EPA Shawnee Test Facility(2), while double-loop venturi/packed towers such as those developed by Research Cottrell can match lime scrubbers in terms of both SO_2 removal efficiency and utilization level(3).

A third factor favoring limestone is the growing realization that forced oxidation can substantially reduce the amount of sludge produced and improve its physical properties for disposal(3,4). Extensive evaluation of forced oxidation at EPA's Shawnee Test Facility has shown that a double-loop scrubber can achieve complete oxidation with either limestone(2,5) or lime(6) feeds when oxidizing at low pH (4.5-5.5) within the first loop. Systems incorporating low-pH oxidation in the scrubbing loop are already being developed commercially(3,7). Tests with EPA's IERL/RTP pilot plant indicate that the conversion to gypsum can also be carried out in a single-loop scrubber operating in the normal pH range of 5.8-6.2 when using limestone feed(8). Considerable simplification of the overall process is thus possible for limestone scrubbers, from both chemical and operational standpoints. Further evaluation of the single-loop approach is undertaken in this paper.

A single-loop scrubber with forced oxidation, illustrated by Figure 1, consists of a SO_2 scrubber and a scrubber-effluent hold tank (EHT) where oxidation is forced by aeration. In the tests reported here, aeration of the EHT was carried out by sparging air into the bottom of the EHT/oxidizer tower, which contained slurry to a depth of 5.5 m (18 ft). Continuous recirculation of the slurry through the SO_2 scrubber keeps the low solubility of $CaSO_3$ from limiting oxidation--the solid dissolves readily in the low pH of the SO_2 scrubber (it would not dissolve in the EHT, which is supersaturated with calcium sulfite, once precipitation occurred). Operation of this system over the past year at IERL/RTP indicates that an EHT/oxidizer will perform well at pH 6, with an air stoichiometry of about 2.5 g atoms oxygen (injected as air) per g mol of SO_2 absorbed in the scrubber.

Oxidation of the slurry in the scrubber loop will obviously affect the system chemistry since $CaSO_3$ is no longer a major participant in the scrubbing reactions. Several important questions are thus raised concerning the possible impact of oxidation on scrubber performance, the results of which are not easily predictable. The purpose of the tests reported here is to answer those questions, which can be set forth specifically as follows:

- The most reliable operation of mist eliminators (with respect to fouling) is obtained when limestone utilization is maintained above 85 percent(1). Can a scrubber operate at this level of utilization when the slurry is oxidized and still attain SO_2 removal efficiencies comparable to those of the unoxidized system?
- The SO_2 removal efficiency of the scrubber is known to depend, in part, on reactions of SO_2 with $SO_3^=$ and $CaSO_3^o$ in the scrubbing liquor(9). If the concentration of these sulfite species is reduced by oxidation, will a significant loss of SO_2 absorption result?

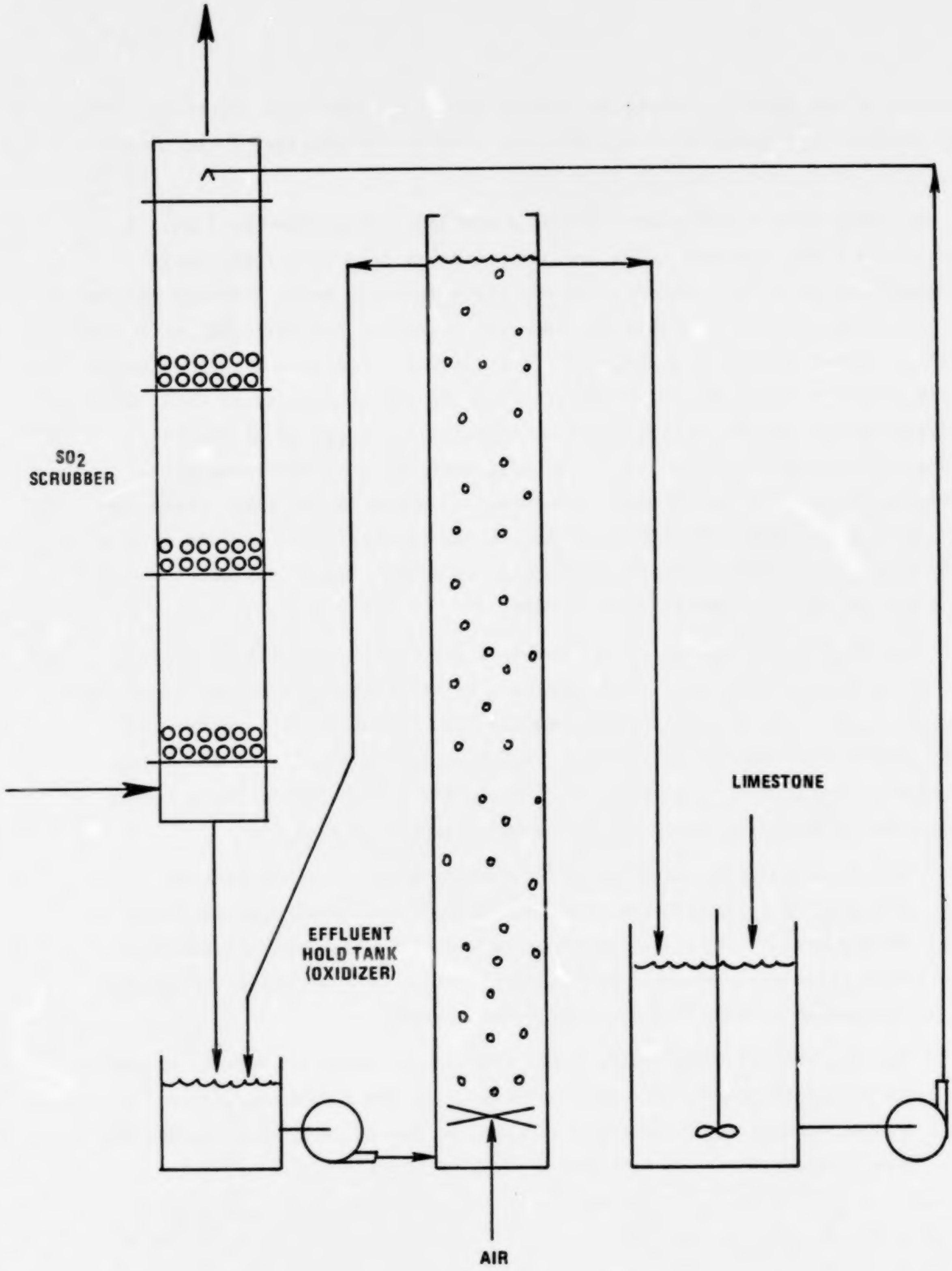

Figure 1. Pilot plant configuration for single-loop forced oxidation tests.

- How will the scrubber feed pH be affected by oxidation?
- The relative supersaturation of $CaSO_4 \cdot 2H_2O$ in the scrubber feed liquor must not exceed 1.4 if scaling is to be avoided. Can this condition be satisfied while oxidizing all of the SO_2?
- Many coals contain chlorine which yields flue gases containing HCl in addition to SO_2. The chloride absorbed by the scrubber accumulates in the scrubbing liquor as $CaCl_2$ to a concentration that is related to the "tightness" of the system; i.e., how much liquor leaves with the sludge. When the sludge is oxidized to improve its dewatering properties, the system will operate with a more tightly closed loop and thus at a much higher level of dissolved chloride. What effect will this chloride have on the SO_2 removal efficiency, limestone utilization, and oxidation efficiency?

These questions are primarily concerned, directly or indirectly, with the effect of oxidation on the SO_2 removal efficiency. The pilot plant tests were therefore conducted to permit comparison of scrubbing efficiency at given operating conditions with and without forced oxidation of the slurry.

PROCEDURE

Tests were conducted over a period of 10 months in a pilot plant containing all components of a complete limestone FGD system. The SO_2 scrubber was a turbulent contact absorber, 23-cm in diam x 3 m high, having a flue gas capacity of 8.5 m^3/min and containing three beds of 3.8-cm diam (5g HDPE) spheres. The bed depth was varied between 18 and 23 cm; the 18-cm beds provided low liquor holdup and low pressure drop while the deeper beds increased holdup and SO_2 removal efficiency. The effect of forced oxidation was thus compared at low and high levels of scrubber holdup.

The scrubber was operated with two different hold tank configurations in an alternating series of runs with and without forced oxidation. The hold tank used for testing without oxidation consisted of a stirred tank containing 718 liters of slurry which was recirculated through the scrubber at a rate of 87 l./min. Forced oxidation was carried out in the scrubber configuration shown in Figure 1; in this case the EHT consisted of a 31-cm diam PVC tower containing 400 liters of slurry at a depth of 5.5 m (18 ft), and a stirred tank containing 318 liters of slurry. The tank in series with the tower provided the same total slurry volume for both testing modes.

The slurry in the oxidation tower was sparged with 6 kg/hr of air by means of PVC pipes extending horizontally across the bottom of the tower. The sparge pipe was 2.5 cm in diam and contained 22 orifices drilled through the pipe wall. The 6.5 mm orifices faced downward to prevent slurry from collecting in, and plugging, the pipe. Air entered the sparger at a pressure of 0.63 kg/cm^2 (9 psig) and was vented from the top of the tower at atmospheric pressure.

The following operating conditions pertain to all tests: scrubber inlet SO_2 conc. = 3000 ppm, flue gas oxygen conc. = 4 percent, slurry temp. = 50°C, and slurry solids = 8 percent (no fly ash was present). The sludge was dewatered in a rotary vacuum filter with the filtrate returned to the scrubber.

Chloride was added by one of two methods: 1) as HCl gas fed into the flue gas entering the scrubber; or 2) as $CaCl_2$ fed with the limestone. The tests with $CaCl_2$ were made to avoid complications in interpreting SO_2 removal efficiencies when HCl is absorbed simultaneously.

A finely ground limestone was used. It was obtained from the EPA Shawnee Test Facility and had the properties indicated in Table 1.

TABLE 1. LIMESTONE USED FOR FORCED OXIDATION TESTS

Particle Size		Composition	
Mesh (Tyler)	Weight Percent	Constituent	Weight Percent
+ 170	0.30	$CaCO_3$	95.0
-170 +230	1.0	$MgCO_3$	1.2
-230 +325	2.5	Insol. Inerts (clay)	2.7
-325	96.2		

The SO_2 removal efficiencies were compared at constant limestone/SO_2 feed ratios and were verified by material balances based on the measured feed rates of SO_2, HCl, and limestone and by the limestone utilizations obtained from analysis of the spent scrubber solids.

RESULTS

Initial comparisons of SO_2 removal efficiencies with and without forced oxidation were made at conditions expected to yield high limestone utilization: deep TCA beds (for maximum slurry holdup), low limestone stoichiometry, and no chloride. The results of these comparisons, summarized in Table 2, show that the scrubber could be operated at limestone utilizations exceeding 85 percent while forcing oxidation. The high utilization needed for maximum operating reliability and minimum operating cost thus appears attainable without loss of SO_2 removal efficiency--relative to the SO_2 removal obtained at the same utilization without oxidation. It is apparent that the oxidized slurry is an effective SO_2 sorbent even at limestone utilizations exceeding 90 percent.

Continuing at low stoichiometry and high scrubber holdup, chloride was fed as HCl with the flue gas. When forced oxidation was not employed, the result of chloride addition was a sharp drop in scrubber feed pH from 6.1 to 5.4 (Table 3). This effect of chloride on pH was expected from experience with the RTP and Shawnee scrubbers and is discussed in more detail below. Although the limestone utilization could be maintained at high levels with chloride present in the system, considerable sacrifice of SO_2 removal was necessary. Reduction of the scrubber holdup to 15 cm H_2O resulted in further loss of SO_2 removal which could be only partially recovered by increasing the limestone stoichiometry (Run C, Table 3).

Forced oxidation was resumed at the same operating conditions used for Run C with respect to the feed rates of limestone, SO_2, and HCl. As indicated by comparing Runs C, D and E of Table 3, two significant results occurred: the pH of the scrubber feed liquor increased from 5.4 to 6.1 and the SO_2 removal efficiency--and limestone utilization--increased.

TABLE 2. SCRUBBER PERFORMANCE WITH HIGH HOLDUP, LOW LIMESTONE STOICHIOMETRY AND NO CHLORIDE

	No Oxidation	Forced Oxidation
SO_2 Removal, %	86	88
Scrubber ΔP, cm H_2O	33	23
Limestone Utilization		
By Solids Analysis	91	97
By Material Balance	90	93
Chloride, ppm	0	0
pH Scrubber Feed	6.0	6.1
pH Scrubber Effluent	5.3	5.2
Oxidation, %	20	98

TABLE 3. SCRUBBER PERFORMANCE WITH CHLORIDE ADDITION[a]

	No Oxidation			Forced Oxidation	
Run	A	B	C	D	E
SO_2 Removal, %	74	67	74	81	79
Scrubber ΔP, cm H_2O	30	15	15	15	15
Limestone Utilization					
By Solids Analysis	95	93	68	89	81
By Material Balance	91	85	64	76	78
Chloride, ppm[b]	19,000	19,000	18,000	16,000	20,000
pH Scrubber Feed	5.4	5.2	5.4	6.1	6.1
pH Scrubber Effluent	4.7	4.5	4.9	5.2	5.2
Oxidation, %	22	17	15	99	99
Filter Cake Solids, %	58	56	54	83	81

[a] Fed as HCl with the flue gas

[b] Concentration in scrubbing liquor

Tables 4 and 5 compare the performance of the system operating with forced oxidation (and high chloride concentration) with the performance obtained at normal chloride levels without forced oxidation. For the latter case, a chloride concentration of about 5000 ppm was chosen, based on an assumed Cl content in the coal of 0.1 wt. percent and dewatering with a settler/clarifier discharging 40 percent solids. The run conditions were selected so that the comparison was made at a limestone utilization of approximately 85 percent when oxidation was forced at the 20,000 ppm chloride level. The difference in chloride concentrations used in the two cases was based on the expected difference in liquor purge rates when the oxidized slurry is filtered to 80 percent solids. The results indicate essentially the same performance in both cases in spite of the higher chloride levels used in the forced oxidation tests.

DISCUSSION

The results of these tests indicate that the benefits of forced oxidation, such as improved sludge quality and reduced waste production, should be attainable in single-loop limestone scrubbers without loss of SO_2 removal efficiency. Performance of a system with forced oxidation in the scrubbing loop should be comparable to its performance without oxidation at any given level of limestone utilization up to at least 90 percent. The pilot plant comparisons indicate that limestone utilizations of 85 percent can be achieved and that forcing the oxidation at these conditions will not reduce utilization--with or without chloride present in the scrubbing liquor.

The effect of forced oxidation on scaling was quite noticeable and beneficial: the scrubber operated scale-free throughout the tests at a $CaSO_4 \cdot 2H_2O$ saturation (about 1.0) far below the critical value for scale formation. The stability of the process with respect to control of sulfate saturation may be one of the more important features of forced oxidation, since saturation is otherwise a function of the uncontrolled variable, oxidation. Oxidation is thus removed from the scrubber tower and transferred to an oxidizer, where it can be controlled. In addition to greatly simplifying the scrubber chemistry, forced oxidation should also improve operating stability and eliminate supersaturation as a variable requiring constant monitoring and control. The

TABLE 4. EFFECT OF FORCED OXIDATION ON PERFORMANCE OF SCRUBBER OPERATING WITH CHLORIDE[a], HIGH SCRUBBER HOLDUP

	No Oxidation	Forced Oxidation
SO_2 Removal, %	91	92
Scrubber ΔP, cm H_2O	36	36
Limestone Utilization		
By Solids Analysis	82	85
By Material Balance	81	83
Chloride, ppm[b]	5,200	22,000
pH Scrubber Feed	5.7	6.1
pH Scrubber Effluent	5.3	5.1
Oxidation, %	27	96
Filter Cake Solids, %	52	79

[a] Fed as $CaCl_2$ to the EHT

[b] Concentration in scrubbing liquor

TABLE 5. EFFECT OF FORCED OXIDATION ON PERFORMANCE OF SCRUBBER OPERATING WITH CHLORIDE[a], LOW SCRUBBER HOLDUP

	No Oxidation	Forced Oxidation
SO_2 Removal, %	84	86
Scrubber ΔP, cm H_2O	24	24
Limestone Utilization		
By Solids Analysis	82	82
By Material Balance	77	84
Chloride, ppm[b]	5,600	30,000
pH Scrubber Feed	5.8	6.1
pH Scrubber Effluent	5.3	5.3
Oxidation, %	15	98
Filter Cake Solids, %	54	80

[a] Fed as $CaCl_2$ to the EHT

[b] Concentration in scrubbing liquor

supersaturation of $CaSO_3 \cdot 1/2\ H_2O$ in the scrubber feed liquor was likewise reduced in the RTP scrubber to half that normally observed without forced oxidation.

From the environmental standpoint, the need to operate scrubbers as closed-loop systems has long been recognized. Increasing the tightness of the loop is also generally regarded as a desirable objective even though the total amount of dissolved pollutants discharged from the scrubber system is not appreciably altered. It is clear from the pilot plant tests reported here that the amount of scrubbing liquor purged to the environment can be substantially reduced by forced oxidation, because of the improved dewatering properties of oxidized sludge. By increasing the solids content of the sludge from 55 to 80 percent, the liquor purge is reduced by a factor of 2.5. When compared with a clarifier/settler discharging 40 percent solids, the reduction amounts to a factor of about 4.

Chloride

Reduction of the liquid purge by any amount requires that the concentration of non-precipitating ionic species increase by a like factor. Calcium chloride is probably the most important of these soluble contaminants. Although this concentrating effect on chloride should enhance the prospects for successful future application of technology for reducing its discharge to the environment (such as evaporation), it is not clear that the performance of the scrubber as an SO_2 absorber can be maintained. The negative effect of chloride, which lowers the steady state pH in closed loop systems and reduces either the SO_2 removal efficiency or limestone utilization, is seen in Table 3 to be virtually eliminated when the hold tank is aerated in the pilot plant. As a result of the higher pH's of the aerated slurry, the negative effect of high chloride concentrations is obviated by forced oxidation. As shown by the tests in Tables 4 and 5, the scrubber performance with oxidized slurry containing up to 30,000 ppm Cl^- was not inferior to that of the unoxidized slurry containing chloride concentrations representative of less-tightly-closed loops producing calcium sulfite sludge.

Limestone Dissolution

The data of Table 6, which compares the average composition of scrubber feed liquors with and without forced oxidation, shows that the concentrations of

TABLE 6. AVERAGE COMPOSITIONS OF SCRUBBER FEED LIQUOR

Constituent	With Forced Oxidation ppm	Without Forced Oxidation ppm
Ca	2280	2300
Mg	660	830
SO_3	70	260
SO_4	1560	1540
CO_3	120	250
Cl	4950	4800
Relative Saturation, $CaSO_3 \cdot 1/2\ H_2O$	= 2.2	5.5
Relative Saturation, $CaSO_4 \cdot 2H_2O$	= 1.0	Variable, 0.2-1.4

dissolved SO_2 and CO_2 are both reduced by about 2 m mol/1. when the EHT is aerated with sufficient intensity to force the oxidation to about 98 percent completion. It is apparent that aeration strips CO_2 from the scrubbing liquor as well as oxidizing SO_2, with an overall reaction that can be represented by:

$$HSO_3^- + 1/2\ O_2 + HCO_3^- \longrightarrow H_2O + CO_2\uparrow + SO_4^= \qquad (1)$$

Without forced oxidation the CO_2 generated by limestone dissolution in the EHT is normally desorbed in the tower where the gas/liquid interfacial area is greatest and the pH is lowest. By forcing oxidation, CO_2 is removed from the liquor before it is returned to the scrubber; it is plausible that the stripping of CO_2 enhances the dissolution of limestone since reaction (1) will also tend to accelerate the reaction:

$$H^+ + CaCO_3 \longrightarrow Ca^{++} + HCO_3^- \qquad (2)$$

by reducing the HCO_3^- concentration.

The tendency of chloride to lower the pH, and the observed effect of aeration to increase it can be explained in terms of reaction (2), since greater concentrations of $CaCl_2$ reduce the solubility of $CaCO_3$ and lower its dissolution rate, while CO_2 stripping has the opposite effect--increasing the solubility of $CaCO_3$ and accelerating its dissolution rate. If the limestone dissolution rate is a controlling factor in the EHT, it will be directly reflected by a corresponding change in the steady state pH. That the dissolution rate is so affected can be demonstrated by laboratory measurements such as those shown in Figures 2 and 3. These data compare the rates of dissolution of Shawnee (Fredonia) limestone at a constant pH of 6.0 in solutions containing zero, 5,000, and 20,000 ppm Cl^- as calcium chloride.* At 20,000 ppm Cl^- the dissolution rate is only one-fourth the rate without chloride. If the rates are measured while sparging the limestone slurry with nitrogen, significantly faster dissolution is observed, as shown in Figure 3, and is no longer affected by calcium chloride. It is postulated that similar effects occur in the EHT of the scrubber during forced oxidation and contribute to the performance of the system.

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*In both experiments 0.25 g limestone (Table 1) was slurried in 200 ml water and titrated with 0.1 N HCl at 30°C.

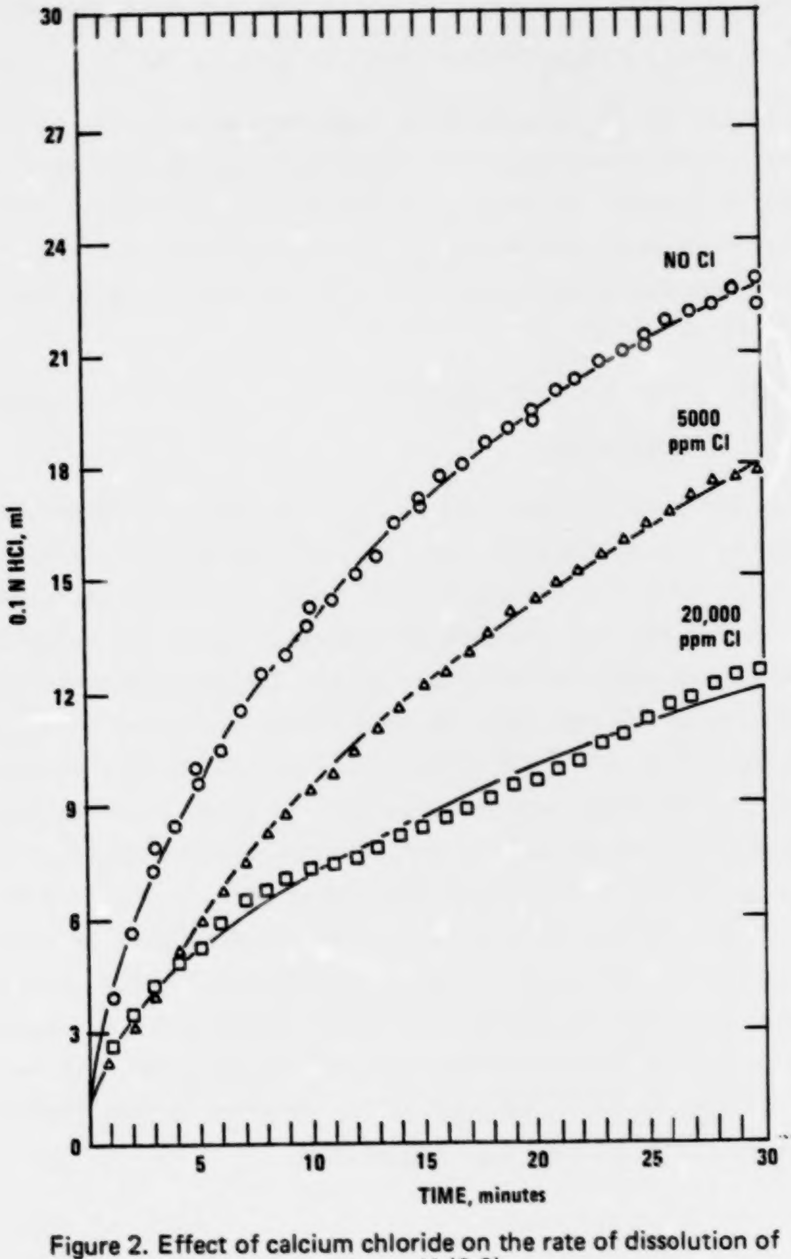

Figure 2. Effect of calcium chloride on the rate of dissolution of Shawnee limestone at constant pH (6.0).

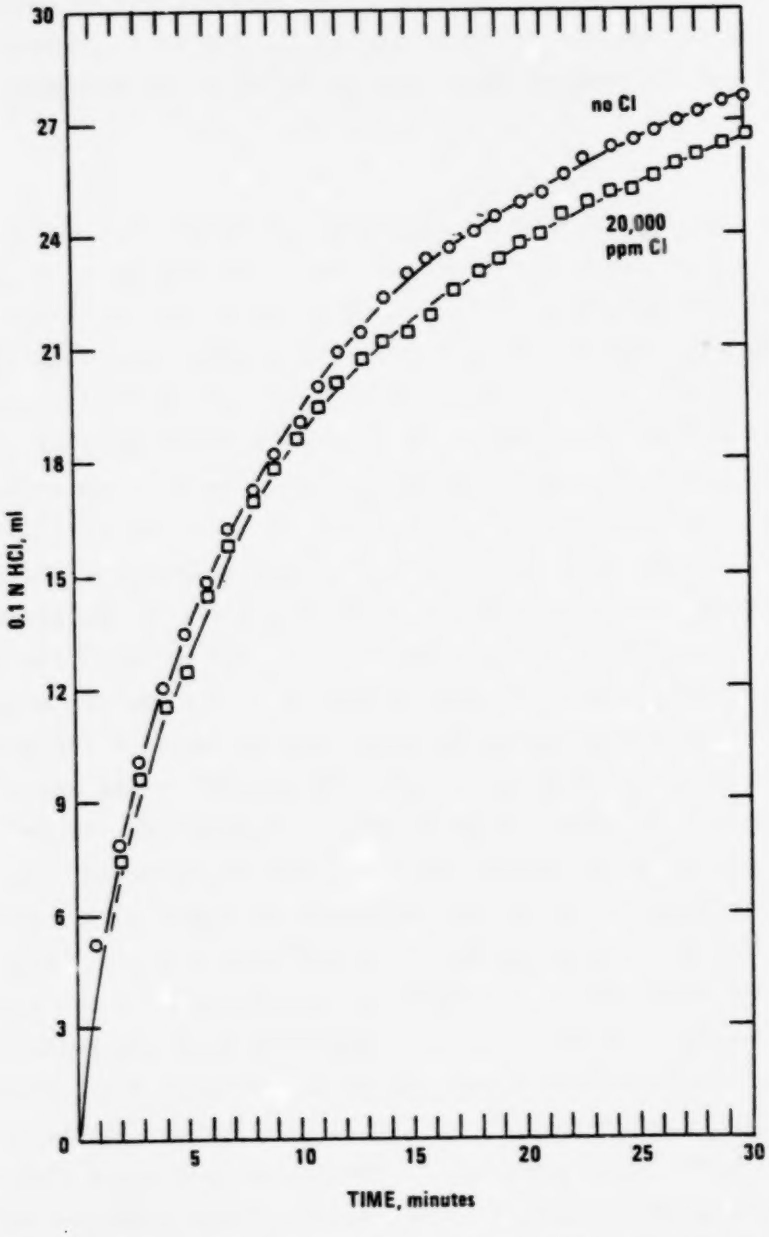

Figure 3. Effect of nitrogen sparging on the dissolution rate of Shawnee limestone at constant pH (6.0).

It is clear from the overall comparisons of SO_2 absorption efficiencies (Tables 2-5) that the removal of solid $CaSO_3$ from the slurry, and the reduction in $SO_3^=$ and $CaSO_3^o$ concentration which result from forced oxidation, do not adversely affect the performance of the scrubber. The capacity for SO_2 absorption that would normally be provided by these species is apparently compensated fully by the higher pH and the reduced HSO_3^- concentration in the scrubbing liquor.

Chloride Control

It has been shown that the application of forced oxidation can improve the dewatering properties of the sludge to the extent that, when filtered, it can be disposed of as landfill (80 percent solids). In addition to the reduction in total sludge production made possible by this approach, another advantage is the potential reduction of soluble salts (mainly calcium chloride) that accumulate in the scrubbing liquor and are eventually released to the environment as sludge leachate. Emerging technology capable of extracting the soluble salts from the scrubbing loop by vapor-compression evaporation(10) or reverse osmosis(11) have been successfully demonstrated. The application of such technology to the SO_x scrubber system can markedly reduce the amount of soluble salts in the sludge while, at the same time, recover pure water for re-use in the boiler. The latter feature can be a crucial one for plants operating in the western states. Taking as a base case an unoxidized system using a clarifier/settler for dewatering, a material balance will show that only about one-fourth the amount of evporation would be required to recover a given amount of pure $CaCl_2$ if the slurry is oxidized and dewatered by filtration. The material balances also show that 70 percent of the chloride could be thus extracted by evaporating 5 percent of the filtrate liquor (or 1.4 liters per kg of SO_2 absorbed in the scrubber), resulting in a 70 percent reduction of the chloride concentration in the scrubbing liquor. As the economics of evaporation are improved by forced oxidation, so also should the scrubber operation be improved by evaporation, e.g., reduced corrosion by chlorides. A mutually beneficial interaction between forced oxidation and evaporation on the performance of the FGD system as a whole thus appears to favor the consideration of both in the design of new limestone scrubbers. A system of this type should make possible the combustion of high chloride, high sulfur coals without damage to the environment.

Oxygen Transfer

The chemical oxidation rate of $CaSO_3$ slurry at 50°C is reported(12) to be 1.4×10^{-3} g mol/l.(min) at pH 6. This rate is quite sufficient to fully oxidize the slurry in a scrubber operating with 3000 ppm inlet SO_2 and a 8-min EHT residence time at L/G = 9.4 $l./m^3$ (70 gal./Mcf) if the full volume of the hold tank is utilized in the oxidation reaction. The use of a narrow deep tank for aeration helps meet this constraint by ensuring good distribution of air bubbles throughout the volume of the oxidizing slurry. Pilot plant experience at RTP shows that an unstirred, air-sparged tank of 5.5 m (18 ft) slurry depth yields good oxidation at pH's up to 6.5; the overall oxidation efficiency in this system appears to be limited by the O_2 transfer from the air bubbles to the slurry. Figure 4 compares the oxygen transfer efficiencies estimated from RTP scrubber tests of air sparged towers(8) with the values predicted by the model developed by Urza and Jackson(13) for O_2 diffusion control. The tests have shown no effect of chloride, positive or negative, upon O_2 transfer in the range of zero to 20,000 ppm Cl^-. The sparger orifice size likewise had no effect from 1.6 to 6.4 mm (1/16 to 1/4 in); the larger orifices are thus recommended to reduce air pressure requirements and reduce plugging. It is reasonable to expect that the same factors which contribute to efficient O_2 transfer will also contribute to good CO_2 stripping; i.e., deep tanks, small air bubbles, and uniform air/liquid distribution.

Although all of the forced oxidation tests made for the purpose of comparing scrubber performance were conducted with the sparged tower configuration shown in Figure 1, good oxidation was also obtained in supplementary tests using a shallow tank of 1.1 m depth and a Penberthy ejector for aeration which produces smaller bubbles than the sparger. Two sizes of ejectors were tested in the 718 l. EHT. The larger model took the entire flow from the scrubber effluent as ejector feed, and yielded 98 percent oxidation at air stoichiometries as low as 1.8 when fed either with recycled slurry from the EHT or with scrubber effluent. The smaller ejector processed only about half of the scrubber effluent flow and, although less effective than the larger model, could also achieve 98 percent oxidation under optimum test conditions at an air stoichiometry of 2.5. The optimum conditions were determined in a

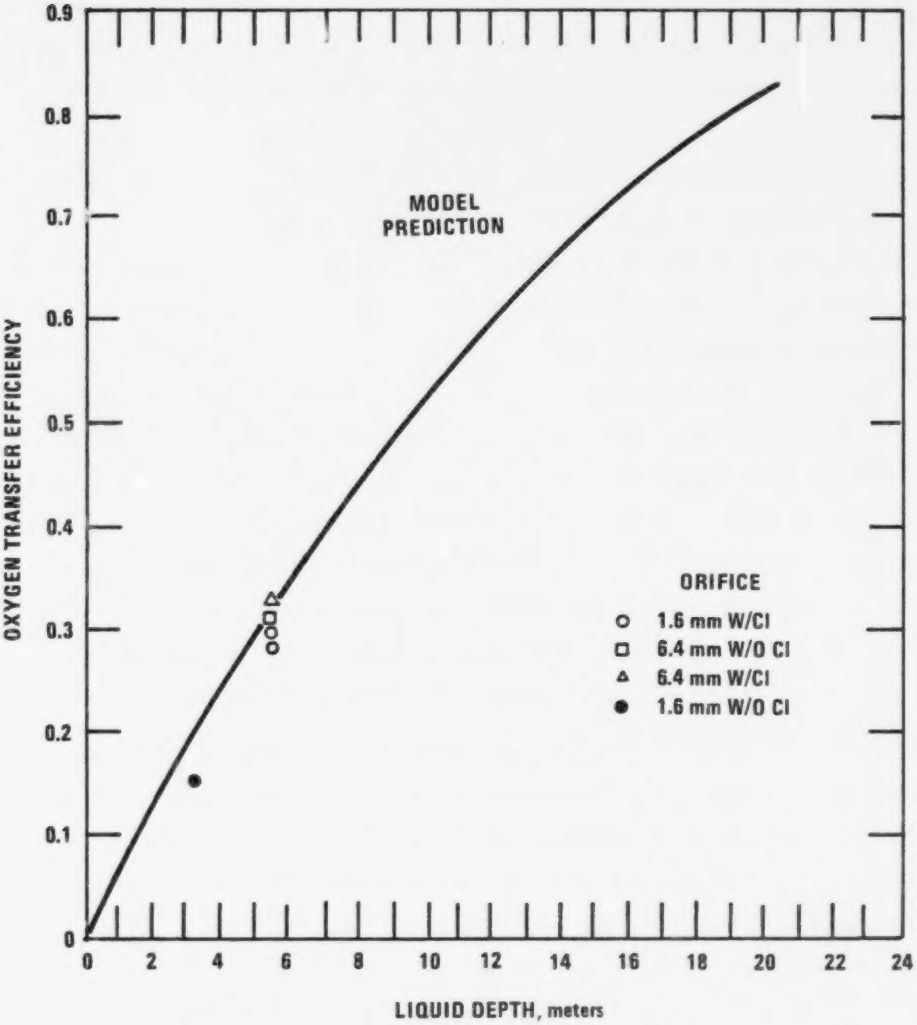

Figure 4. Comparison of O_2 transfer efficiencies obtained in the RTP pilot plant with the liquid-film diffusion model at 50° C.

series of tests with the small ejector which are summarized in Table 7. The best results were obtained when the ejector was mounted in the side of the EHT, near the bottom, and fed with slurry taken from the scrubber effluent. Oxidation was incomplete when the slurry feed was switched to the EHT. Performance was poorest when the ejector was mounted vertically on top of the hold tank (ejecting down into the EHT slurry).

CONCLUSIONS

Summarizing EPA's IERL/RTP experience with the application of forced oxidation to a single-loop scrubber operating at 3000 ppm inlet SO_2:

- Forced oxidation could be carried out in a single-loop limestone scrubber operating at normal pH.
- The SO_2 removal efficiency was not adversely affected by oxidizing the slurry in the scrubbing loop, whether or not chloride was present.
- With good aeration of the EHT (and finely ground limestone) the scrubber feed pH was about 6.1 regardless of the chloride content of the scrubbing liquor.
- The scrubber could operate at 85 percent limestone utilization with SO_2 removal efficiencies comparable to those obtained without oxidation; increasing the scrubber holdup enhanced SO_2 removal at any given level of limestone utilization.
- The scrubber feed supersaturation was well below the critical value for sulfate scaling and was more stable than that normally observed without forced oxidation.
- An EHT residence time of 8 min, relative to scrubber recycle at $L/G = 9.4$ l./m^3 (70 gal./Mcf), was sufficient to produce good sludge properties, similar to that obtained with a double-loop system at low pH. The slurry was consistently filterable to 80 percent solids.
- An EHT/oxidizer of 5.5 m (18 ft) depth provided adequate O_2 transfer at pH 6.1 to oxidize the slurry with 2.5 g atoms O_2 (injected as air) per g mol of SO_2 absorbed when aerated with a sparger containing 6.4 mm (0.25 in.) orifices.

TABLE 7. PERFORMANCE OF PENBERTHY EJECTOR AT 48 l./min SLURRY FLOW RATE

Test	Test Mode	Ejector Feed[a]	Percent Oxidation	Filter Cake % Solids	Settling Rate, cm/min
1	Horizontal Ejector (Tank Bottom)	HT	49	55	0.2
2	Horizontal Ejector (Tank Bottom)	SE	99	81	2.8
3	Vertical Ejector (Tank Top)	SE	67	62	0.4
4	Vertical Ejector (Tank Top)	SE	53	58	0.4
5	Horizontal Ejector (Tank Bottom)	SE	98	86	3.5
6	Horizontal Ejector (Tank Bottom)	HT	87	65	0.7

[a] HT = ejector feed taken from EHT; SE = ejector feed taken from scrubber effluent

- Ejector oxidation was most efficient when the ejector was fed with scrubber effluent slurry and mounted horizontally in the side of the EHT, near the bottom.

- The performance of the oxidizer was not affected by chloride or limestone stoichiometry.

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OPERATING EXPERIENCE, BRUCE MANSFIELD PLANT FLUE GAS DESULFURIZATION SYSTEM

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ABSTRACT

This paper describes the Bruce Mansfield Plant and the flue gas desulfurization system associated with the plant. It briefly discusses the preliminary studies made to determine the type of scrubbing system to be used, describes in detail the system finally installed, and discusses many of the problems involved in the startup and operation of the scrubbing system, along with scrubber chemistry and mist eliminator operation.

Operating Experience
Bruce Mansfield Plant
FLUE GAS DESULFURIZATION SYSTEM

The Bruce Mansfield Plant is located in the southwestern part of Pennsylvania on the Ohio River in Shippingport, Pennsylvania. The plant is being constructed and will be operated by Pennsylvania Power Company, a subsidiary of Ohio Edison Company. It is owned by the Central Area Power Coordination Group (CAPCO) made up of Ohio Edison Company, Pennsylvania Power Company, the Cleveland Electric Illuminating Company, Duquesne Light Company, and the Toledo Edison Company. The original concept included the installation of two 880,000 (net) KW coal-fired generating units of conventional design with steam conditions of 3500 psi, 1000 DegF main steam and 1000 DegF reheat. The air quality control system envisioned at that time incorporated high-efficiency electrostatic precipitators and a tall chimney. Engineering for the project started late in 1969 and construction started in June, 1971.

At the start of engineering the Commonwealth of Pennsylvania did not have a state-wide standard applicable to sulfur emissions. Engineering proceeded on the basis of Pennsylvania Regulation V (since superseded by Chapter 123.11) which established a very stringent particulate emission limitation. The Company was aware that SO_2 emission regulations might be promulgated sometime in the future, and, therefore, engineering effort was directed toward the installation of an air quality control system satisfying the particulate emission regulation that could be later coupled to an SO_2 removal device of some kind. This prompted initiating an investigation during early 1970 of the state-of-the-art and the expected commercial availability of SO_2 removal systems.

When engineering design had proceeded to the point at which the Company could notify the Pennsylvania Department of Environmental Resources (DER) of the design concept and sizing of major equipment, a partially completed form for Application for Construction Approval was submitted. In November of 1970 the Pennsylvania DER advised the Company that they had reviewed the preliminary Application and had concluded that since the design did not include a full-scale SO_2 removal system, it was unlikely that approval of the Application would be recommended, even though Pennsylvania apparently had no applicable SO_2 emission limitation at that time. Nevertheless, this notification accelerated and intensified a series of studies then in progress in connection with SO_2 removal systems.

The studies concluded that the technology associated with SO_2 removal systems was then in an early state of development and totally unproven. For this reason, the Company proposed discussions with the Pennsylvania DER and the Federal Environmental Protection Agency (EPA) to build a module of 20% to 25% size to serve #1 unit. This module would provide the vehicle to solve operating, chemical, and disposal problems. It would also prevent severe operating problems and outages of equipment required to make modifications on the full number of modules in a system installed to treat all of the boiler flue gas and would allow the remaining portions of the SO_2 removal modules not initially installed to be successfully operated at an earlier date. In the interim period a tall chimney (800' to 1000') would prevent ground level SO_2 concentrations from exceeding the ambient air quality standards. (A 950' chimney height was later recommended pursuant to a study performed by Battelle Columbus Laboratories.) It was further pointed out that the single module approach would avoid the cost of duplicating modifications, which no doubt operating experience would prove necessary.

As a part of this study a world-wide search was conducted for vendors capable of supplying a full-scale SO_2 removal system of the size required on the Mansfield units with a performance warranty that would satisfy the requirements of the Pennsylvania SO_2 emission regulations. Only one vendor was willing to undertake this full-scale project, and that vendor limited the performance warranty to a very modest amount. The DER and EPA did not approve the application for a 20% to 25% module for Unit #1, and, therefore, a full-scale SO_2 removal system was constructed. Total cost for the first two units is approximately $750 million; included in this $750 million is about $240 million for the cost of the air quality control system.

The study concluded that the only arrangement showing any promise of achieving the level of SO_2 removal efficiency required by the Regulations, about 92%, was the installation of several trains of two-stage wet lime scrubbing. Six trains per unit were chosen in order to minimize the magnitude of scale-up in size and to provide a spare train for each unit.

Pennsylvania Power Company also commissioned a parallel study with a company to determine and evaluate problems involved in the disposal of wastes produced by the lime throw-away system. After very detailed studies of several methods of disposal, it was concluded that the only one available that provided reasonable assurance of continued operation of the plant was the treatment of the sludge with a compound then under development that would tend to cause it to solidify over some period of time after pumping it to an impounded disposal area.

The large quantities of sludge generated, expected to be upwards of 2,000,000 tons per year from the two units, require a very substantial land area for disposal. To satisfy this need, some 1400 acres of property was purchased (an entire valley) about six miles from the plant site. A 420' high hydraulic dam was constructed to contain the sludge in this area.

The two-stage venturi scrubber for Mansfield Units 1 & 2 is designed and furnished by Chemico. As indicated above, six trains are installed, each train consisting of a scrubber vessel, induced draft fan, and an absorber vessel. There are two oil-fired reheaters. Three scrubber trains discharge into a single reheater. The scrubber and absorber vessels are about 35' in diameter and 50' high. The vessels and ductwork are lined with polyester flakeglass material. The induced draft fan housing is lined with rubber, and the induced draft fan rotors are made of Inconel-625.

The flue gas enters the top of the scrubber vessel, passes down and around an adjustable plumbob through the venturi throat, turns 180°, passes up through a mist eliminator out of the absorber vessel and into the suction of the induced draft fan. The adjustable plumbob makes it possible to control the velocity of the gas at the throat of the absorber. Pressure drop across the throat is maintained at approximately 20" water. Scrubber liquor is circulated with two recycle pumps taking their suction from the base of the scrubber vessel, and then pumping the liquor to the top of the vessel where the venturi throat and plumbob surfaces are wetted. The pumps also supply liquor to secondary sprays located in the gas stream ahead of the mist eliminator. An intimate mixing of the gas and liquor in the venturi throat and with the secondary sprays is designed to remove practically all of the particulate matter and about 70% of the sulfur dioxide.

From the induced draft fan the gas enters the top of the absorber vessel, passes down through a fixed throat venturi, turns 180° and passes through a mist eliminator, leaves the absorber vessel, and enters the reheater. Absorber liquor is circulated with two recycle pumps in a manner similar to the scrubber. The absorber is designed to remove the remaining particulate matter and enough sulfur dioxide to meet the Commonwealth of Pennsylvania emission standards. The water-saturated gas enters the reheater at about 125 DegF and is designed to leave the reheater at about 165 DegF, thus evaporating any moisture carry-over, adding buoyancy to the gas and preventing precipitation from the stack plume in the vicinity of the plant. The pH of the scrubber and absorber liquor is maintained between 7 and 8 by the addition of lime to these vessels.

Enough of the liquor in the absorber is bled from the discharge of the recycle pumps to the scrubber vessel to maintain 8-10% solids in the absorber liquor. The scrubber vessel liquor is bled in the same manner but to a 200' diameter thickener located on the plant site. Thickener overflow water is used to maintain a liquid level in the scrubber and absorber vessels. Thickener underflow is controlled to approximately 30% solids and pumped to the 1400 acre impoundment site with large reciprocating sludge pumps.

The pumping station is located on the plant site and pumps the sludge over six miles to the impoundment area without booster pumps. The Dravo Corporation designed and built the sludge handling system including the hydraulic dam for the impoundment area. Dravo also supplies the lime and the sludge hardening agent called calcilox for this project. Calcilox is added to the sludge on the plant site at the sludge pump suction. This material is designed to solidify the sludge to the extent that it will support about four and one-half tons per square foot or equal to solid earth. Operating costs are high. The air quality control system almost doubles the station power requirements, and along with the circulating water cooling tower, reduces the net output from an original design of 880 MW to 825 MW.

Unit #1 was placed on the line December 11, 1975, and was placed in full commercial operation June 1, 1976. Unit #2 was placed in full commercial operation October 1, 1977. Operating experience with the scrubbers has exposed certain problems requiring modifications to the system. The cavitation of the recycle pumps caused pulsing of the liquor flow and, therefore, an unstable gas flow through the scrubber vessel. Baffle plates were installed over the pump suction openings in the scrubber vessel which solved the problem but required several modifications of the baffles before one was found that could withstand the forces exerted upon it and the corrosive atmosphere that it was exposed to.

One of the scrubber mist eliminators was plugged up very early in the operation of the system. The material plugging the mist eliminator hardened to a point where it could not be removed, and the mist eliminator had to be replaced. The mist eliminators are normally cleaned by intermittent spraying with recycle and fresh water. After the problem of the mist eliminator pluggage occurred, a system was installed that permitted the mist eliminator to be flooded with large volumes of water in case the pressure drop across the device became excessive.

Tests conducted about a year ago indicated that carryover from the mist eliminators was on the order of two to three grains per cubic foot, while design for this system is a maximum of one grain per cubic foot. On the basis of these tests and an obvious problem of precipitation from the stack plume, it was decided to experiment with a vertical mist eliminator installed in the outlet duct of the absorber vessel.

After some difficulty, a vertical mist eliminator was installed and successfully operated in this area on one scrubber train. In June of this year Chemico had some model studies conducted on both the horizontal and vertical mist eliminators. Full-size mist eliminator sections were used in the model studies, and the information gained from the study was very helpful to the operating company in establishing a certain criteria relative to the operation of the mist eliminators. It was found in the model study that pressure drops in excess of three-quarters to one inch water allowed excessive carryover from the horizontal mist eliminators. When the pressure drops were maintained at half inch water or less, there was practically no carryover of entrained water.

It was obvious after seeing the model tests that it was essential to operate the horizontal mist eliminators with low pressure drops. If the pressure drop exceeds .3" H_2O, action is taken to clean the mist eliminator. If scale forms on the mist eliminators and the pressure drop tends to increase, then the scrubber train has to be shutdown and the mist eliminators cleaned in some manner, or while the scrubber train is in service the mist eliminators can be flooded with fresh water and the scale can sometimes be washed off. The ideal situation, of course, is to maintain a very strict chemical control over the scrubber cycle and, therefore, eliminate the formation of scale on the mist eliminators.

If the pH of the scrubber liquor is maintained between 7 and 8, and if the magnesium dissolved in the scrubber liquor is maintained at about 1500 ppm or more, then the scale will not form on the mist eliminators, and scale is less of a problem in other parts of the system. It is also important to keep the gas flow through the system at or below design; otherwise, scrubbing liquor is physically carried into the mist eliminators and causes scaling. Measuring flow is difficult because there are no long straight duct runs. Originally, it was felt that pressure drop across the fixed throat of the absorber would represent flow, but this has not proven to be a reliable flow-measuring device. A pitot tube flow-measuring system has been purchased and is in the process of being installed at the scrubber inlet.

Flooding the mist eliminators with fresh water while the unit is in operation dilutes the liquor in the scrubber vessel and causes the magnesium content to decrease, which in turn causes the scaling properties to increase, and the SO_2 removal efficiency to decrease. Operation of the mist eliminators is very closely associated with the chemistry of the system, and the key to maintaining the proper chemistry is (1) accurately measuring pH continuously, and (2) maintaining a high concentration of dissolved magnesium in the scrubber liquid.

The original design for measuring pH in each of the scrubber and absorber vessels was intended to record and indicate the pH to the operator in the air quality control system control room and for automatic control of lime feed into each vessel. These are flow-through cells, and the difficulties were primarily breakage of the glass measuring electrode and maintaining a sample flow through the system. The equipment was also flow sensitive, and for accurate pH measure flow must be maintained at less than 15 gpm. After many months of modification and experimentation, the pH measuring equipment was moved to a more desirable location in the scrubber cycle and appears to be working satisfactorily. Maintenance of this equipment is and will continue to be high.

The major operating problem associated with the Bruce Mansfield Plant flue gas desulfurization system seems to have been centered around pH control. Contributing factors have been the inability to measure flue gas flow through the scrubber trains, a lack of understanding of the mist eliminators and their operating limitations, and problems with mechanical equipment.

Air quality control system operating less maintenance costs for Unit #1 for the first eight months of 1977 amounted to 3.83 mills per KWH. Maintenance costs for the same period was 1.65 mills per KWH. Total plant operating costs for the same period was 19.5 mills per KWH. Since Unit #2 was placed in commercial operation on October 1, 1977, meaningful cost data is not available for that unit. Hopefully, the costs quoted for Unit #1 will not be typical for the life of the unit because the time period includes a ten-week maintenance outage and a long period of partial load due to a chimney flue liner failure.

The unit has been at half load since returning to service after the ten-week outage in May because the polyester flakeglass lining in the chimney flues in Unit #1 failed and caused extensive damage to the carbon steel flue liners. There are two of these flues; they can be repaired one at a time with the Unit operating at approximately half load. One flue has been repaired and the second one is in the process of being repaired. It is expected that the Unit will be back to full load by January, 1978. A very similar polyester flakeglass lining has held up reasonably well in the scrubber and absorber vessels and the ductwork. We are presently experimenting with test patches of various linings in one flue of Unit #2. Currently, we do not know of any lining that we would have confidence in.

The Pennsylvania DER conducted stack tests in July and found that particulate removal for the Bruce Mansfield system is excellent. In one of the tests for sulfur dioxide, the stack emission met the State standards. In the second test the SO_2 emission was higher than the State standards of .6 lbs/million BTU input because of some problems with the lime feed system. Since better control of the scrubbing system pH has been achieved, SO_2 removal efficiency has been consistently high enough to meet the State standards. This consistency has only been achieved in the past several weeks, and whether or not it will continue to operate satisfactorily is still in question.

Several major problems still exist. The air quality control system was designed to operate with a closed-loop water system, but this has never been achieved. As mentioned above, the chimney liners are still in question. Reheat burners as they are presently installed will never work properly. Mist eliminators may still be a problem, and there are many problems of a minor nature that remain to be solved.

LOUISVILLE GAS AND ELECTRIC COMPANY SCRUBBER EXPERIENCES AND PLANS

Robert P. Van Ness
Louisville Gas & Electric Company

ABSTRACT

This presentation has been divided into two sections. Section I contains an up-to-date report of the progress on the installation of four carbide lime wet scrubbing SO_2 removal systems on electric generating Units #5 and #6 at Cane Run Station and Units #3 and #4 at Mill Creek Station, and a discussion of the operating experience on the system installed on Unit #4 at Cane Run Station, which went in service on August 7, 1976. The progressive installation of these systems is in conformance with a consent agreement negotiated with the U.S. Environmental Protection Agency, Region IV and the Air Pollution Control District of Jefferson County, Kentucky.

Section II contains a preliminary report of the progress on the research project under a grant from the Environmental Protection Agency. The project includes: (1) a test program of the operation of the 65-MW Paddy's Run Unit #6 marble bed wet scrubber SO_2 removal system under varied operating conditions and with the use of alternate slurry additives (2) studies, integrated with the operating test program, of various methods for the disposal of scrubber sludge in an environmentally acceptable manner.

Louisville Gas and Electric Company
Scrubber Experiences and Plans

SECTION I

During the years 1974 and 1975, the Louisville Gas & Electric Company had many discussions and meetings with the Environmental Protection Agency and the Air Pollution Control District of Jefferson County in an attempt to establish a compliance plan for sulfur dioxide control at all of our facilities in Jefferson County. These discussions culminated in consent decree signed on December 10, 1975 in which sulfur dioxide removal systems were to be installed on various boiler units at our generation stations at Cane Run and Mill Creek. This enforcement order in effect mandated the installation of seven SO_2 systems which includes Cane Run Nos. 4, 5 & 6 and Mill Creek Nos. 1, 2, 3 & 4. In a revised order of September 1, 1977 this order recognizes various delays in the construction schedules on these systems outlined in the original order of 1975 due to equipment delivery delays and other construction difficulties beyond our control.

The first commercial unit which our construction crews installed went into operation on August 7, 1976. This system was supplied by the American Air Filter Company and consisted of two mobile bed contactors on Cane Run Unit No. 4 (178 MW). This unit, as well as all other units in our system, operate with the additive known as carbide lime and is designed to remove at least 85% of the SO_2 from the flue gas when burning approximately a four percent sulfur coal. After many difficulties and various modifications, to be discussed later, the unit came into compliance after Performance Testing by EPA on August 3 and 4, 1977. Since that date, the unit has operated quite well with only minor difficulties at maximum load.

The second unit under the compliance plan is Cane Run Unit No. 5 (183 MW) which is presently nearing completion. This unit supplied by Combustion Engineering Company of Windsor, Conn., is scheduled to go into operation during December of this year. This system is a two module spray tower system with steam reheat. The system contains a common reaction tank under the two modules and a thickener for sludge clarification. Performance Testing of the unit should be completed by March 1, 1978.

The third unit mandated by the order and which is presently under construction is another American Air Filter unit with four mobile bed spray towers on a new B&W boiler installation with electrostatic precipitators by AAF. This 425 MW unit should go into operation on or about January 1, 1978 and the scrubber system Performance Testing should occur before April 1, 1978. This system was redesigned to incorporate the modifications required on Cane Run No. 4. Steam reheat of the exit gases from the scrubber system is also being supplied. The scrubber linings for this entire SO_2 system will be precrete over the low carbon steel plate.

A fourth unit of the dual alkali concept is now under construction on Cane Run Unit No. 6 (278 MW) which should be completed by November 1, 1978. This system is being supplied by Combustion Equipment Associates, New York, N.Y. and will be a carbide lime — soda ash system. The entire cost of this installation of $16,300,000 is being borne by LG&E, however, performance testing and a year's operational study is being funded by an EPA grant of $4,500,000. This testing program should last about fifteen months. The EPA support contractor is the Betchel Corp. of San Francisco, California. The Project Officer for EPA will be Norm Kaplan and I will be the Project Director. Sulfur dioxide removal efficiencies of greater than 95% are guaranteed by CEA as well as an electrical loading not to exceed 1.2 percent of the boiler output. Restrictive use of soda ash to 0.045 moles per mole of sulfur removed and a filtered sludge cake of 55% are also required. Performance Testing of the scrubber unit should occur before February 1, 1979. The duct tie-ins to the boiler and its by-pass provisions commenced on October 24.

The last mandated scrubber system is being installed on Mill Creek Unit No. 4 (495 MW) and is being supplied by AAF. The system, including the boiler (B&W), electrostatic precipitators (AAF) and the cooling tower (Zurn) is very similar to Mill Creek No. 3. This new unit is scheduled for start-up by July 1, 1980. This scrubber unit will also be a four module mobile bed spray tower system. Steam reheat and by-pass of individual modules are the same as on Mill Creek No. 3 and Cane Run No. 4.

As previously stated, Cane Run Unit No. 4 ran into serious operating difficulties a week or two after start-up in August 1976. The main initial problem occurred with excessive pressure drop across the system thus restricting the unit to about 150 MW. This pressure drop problem was analyzed to be in the duct work system, the rotary demister and the quenchers. The SO_2 removal efficiencies, due to malfunctions in the spray nozzle system and the spray pump system, were less than satisfactory. Various modifications commenced in early September and continued intermittently throughout the rest of the year with fairly good operability data of about 90% on the system. Various modifications during this period included the installation of new ceramic spray nozzles, installation of turning vanes out of the quench sections and into the contactor vessels, cutting away sections of the demister wheels and various other minor modifications. These modifications appeared to improve overall operability up to maximum load conditions of 190 MW, but at full load operations the system produced fairly poor SO_2 removal efficiencies of between 75 and 80 percent. So called plugging and scaling problems, however, have never been a problem.

Obviously, with these fairly unsatisfactory results on SO_2 removal rates further modifications were needed. I must add here that American Air Filter Company cooperated with our suggestion for a basic redesign of the system to increase the l/g requirement, to improve the demisting quality of the system, to replace the failing Carboline lining in the demister cans and the outlet ducts to the stack, and to add reheat to the system. AAF modeled the SO_2 improvements in their pilot plant and came up with various suggestions which ultimately were incorporated during a major shutdown this spring and early summer. These basic decisions were made in early January of this year and the shutdown was scheduled for mid-April after all the tests were completed and the equipment was on the job site. As the winter came on in full force in early January, the scrubber system, though operative, ran out of carbide lime slurry due to the freezing of the Ohio River that curtailed the lime slurry barge delivery until mid-March. During this period of sub-zero weather, the scrubber system was operated in a slurry-recycling mode without flue gas so that the associated piping system would not freeze-up. This proved quite successful and no apparent problems of serious freezing conditions were encountered; however, we did warm the system up for a period of four hours every two weeks. When the lime slurry system was reestablished in early March the system was operated in various tests modes with quite good operability until shutdown on April 18, 1977, along with the boiler, for a projected two-month overhaul period.

During the major modification period, the following changes and improvements were made in the entire system:

1. Removed all the acid brickwork in the 250 foot concrete stack and relined the stack with a precrete lining. As of this date this lining has held up very well.
2. Effectively removed the demister wheel on each scrubber, by cutting 18 inch holes in the top section of the wheel, and replaced this demisting section with two banks of chevron demisters and its associated spray washing system. Since start-up in mid-July, the demisters have operated quite well and have remained very clean to this date.
3. Installed oil reheat burners in the exit duct work as it enters the stack and installed turning vanes at this point to provide good mixing.
4. Installed a new spray header system above the original mobile bed spray header basically to improve the gas distribution and increase the liquid to gas ratio. This has provided a superior gas contacting mechanism which basically has increased the SO_2 removal efficiency above the 85% legal requirement.
5. Replaced the badly blistered Carboline lining in the demister sections and outlet ducts with Plasite 4005. As of this date, the lining is holding up very well. The lengthy installation of this material delayed the start-up of system almost a month beyond the projected two-month shutdown.
6. Various minor modifications too numerous to mention here.

Since the start-up of the system again on July 17, 1977, the system has operated very well and has consistently been in compliance. One major problem, unrelated to scrubber operability or removal efficiency revolves around the guillotine gate system which has been extremely troublesome. Namely, and still unresolved at this time, is the inability of the individual gates to track smoothly without excessive sticking. Further modifications will be necessary at next shutdown.

As of this date, the system appears to be a satisfactory sulfur dioxide removal system at all levels of boiler load. Availability and reliability have consistently been good over the last few months, with data well above the 90% level.

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SECTION II

The remaining portion of this paper, in a slightly abbreviated form, will cover the preliminary results and findings, to date, of the scrubber and sludge study at Paddy's Run No. 6 marble bed scrubber under a $1,800,000 EPA grant. The actual test program itself consisted of four phases as shown below.

Phase I - Carbide Lime Characterizations and Sludge Mixing Program.
Phase II - Commercial Lime Testing and Sludge Mixing Program.
Phase III - Hold Tank Modifications
Phase IV - Mg/Cl Additions

In Phase I, the carbide lime testing went on quite uneventfully and reestablished to a degree the findings we have reported in the past. The gypsum relative saturation as expected, remained subsaturated in most of the tests run with a feed saturation of between .7 and .75 while the outlet approached a range of .9 to 1.0. Oxidation rates in the solids were between 7 and 8 percent. During the testing program, a contaminate in carbide lime process water, which we name Specie "x", tended to foul-up the chemical analysis work, especially when endeavoring to determine sulfite ion concentrations. As of this date, Specie "x" has not been identified and its effect on the overall system is unknown.

In Phase II, work on commercial lime (Miss. Lime), the same lime used to manufacture calcium carbide and thus its ultimate by-product, carbide lime, produced results in this marble bed scrubber system at the outset which were quite different than those experienced with carbide lime. The system operated in or close to a super-saturated mode most of the time and produced large masses of gypsum under varying control conditions. Increasing liquid to gas ratios and increasing system solids in the slurry system had very little effect on relative saturation levels. Basically, gypsum relative saturation in the feed liquid averaged between .9 to 1.0 and the exit liquid ranged from 1.1 to 1.6. Oxidation increased to 13 to 15 percent in the solids. Sulfur dioxide removal rates, however, were not affected one way or another with respect to a comparison with carbide lime data. An interesting phenomenon occurred at the start of this commercial lime testing phase; the reappearance of Specie "x". In tracking this down, it was found that Airco, in slaking the commercial lime, used the slaking water from the clear overflow from the carbide lime operational thickener while producing carbide lime slurry for use at Cane Run No. 4. By eliminating this water and converting the slaking operation to a city water supply, Specie "x" disappeared from the scrubber slurry system, but did not materially change the gypsum relative saturations in the scrubber circuit. The only conclusion being that, if an antioxidant does occur in carbide lime, it does not exist in the liquid phase.

After running the program for several weeks with the system bordering on a scaling condition, it was shut down to await a decision by all parties, including EPA, as to what procedure should be followed from that point on.

The basic decision ar[illegible] at by all parties was a clean-up of the scale condition in the scrubber with a carbide lime slurry and then phasing into the magnesium additive test program with commercial lime. This was accomplished without difficulty by running the system sub-saturated with carbide lime and dissolving the gypsum concentrations. After a few days of operation in this mode, the scrubber system returned to an extremely clean state. During this operation, a form of carbide lime (black lime) was used that contained varying amounts of magnesium hydroxide. The test data appears to indicate a liquid phase concentration of magnesium in the range of 1000 to 1600 ppm. SO_2 removal rates during the period were in the very high 90% range. This is the same data experienced before while operating under this type of operation during the years 1974, 1975 and 1976.

In actuality, Phase III and Phase IV testing work was somewhat interchangeable; however, we phased into the commercial lime with an addition of a 55% slurry of magnesium hydroxide from Dow Chemical which maintained the liquid phase magnesium in the 4000 ppm range. After a weeks time, this was gradually lowered to the 2000 ppm level. The interesting data secured was a very high SO_2 removal rate with gypsum relative saturations approaching zero. As we approached 2000 ppm during the latter days of June, calcium values skyrocketed as well as relative saturation levels. This was corrected quickly by simply inoculating the system with a little more magnesium. In less than a day's time, the system returned to its extremely low saturation level and remained there for the most of the month of July while the sludge mixing program was in progress. All during this period, extremely high SO_2 removal rates were experienced. The best control range of magnesium for the double marble bed scrubber, using Miss. Lime as the main calcium additive, was between 2400 ppm and 3000 ppm. Additional data should be forth coming next year when we complete evaluation of the sludge mix program and issue the final report.

During the month of August, two other tests were accomplished in that we evaluated the effect of completely eliminating our reaction tank and reducing the reaction time from about 35 minutes to a 3 to 5 minute level. Basically, the above conditions were duplicated with no significant problems, particularly with regard to sulfate scaling. This reduction in residence time might have resulted in calcium sulfite scaling although due to the short test duration no attempt was made to determine whether sulfite scaling did occur. Also the thickener underflow solids data indicates that the solids content deteriorated with the reduced reaction tank residence time.

In many cases, reaction tank design can be a rather expensive cost element in the ultimate design of a scrubber system, especially when considering pump and pipe design. Therefore, the above factors should be considered in designing reaction tanks.

The next phase of testing, which concluded the overall testing program on August 31, 1977, was the inoculation of calcium chloride into the system, with the same short reaction time mode, to assimulate a high chloride type coal. The initial inoculation was at the 9000 ppm level and then was reduced to the 3000 ppm level. To compensate for this addition, magnesium levels were increased to the 3500 ppm level. The result: high SO_2 removal rates in the 99% area and very low gypsum relative saturation levels. No operational problems resulted from operating in this type of mode. The conclusion being, though not run for any great period of time, that high levels of chlorides can be controlled without difficulties by the counter balancing effect of relatively high values of magnesium.

All in all, the overall test program in all its phases was quite valuable and provided some very interesting results. All the answers were not established; however, a better understanding of the scrubber mechanism was achieved, at least by yours truly. This type of data and results have to have some very meaningful significance for future scrubber design.

With respect to the sludge mix program, all the test results, including leachate data, are not in as yet and may not be available until next year; however, the status of the program seems to indicate the results will be quite meaningful. The following summary is, to a degree, the summation of some of the findings of the field sampling and testing program conducted by the University of Louisville and Combustion Engineering on the FGD sludge impoundments, which include ten swimming pools and five pit impoundments at the Cane Run Station.

In Exhibit "A" is a tabulation of the entire sludge impoundment site with the formulations of various mixes and the percent solids of the resulting mixes as they were placed in the impoundments. As a matter of fact, Pools No. 1 thru No. 4 and Pits 1 and 2 were carbide lime sludges that were mixed in October and November 1976, while the remaining impoundments were placed in July 1977 with a mix that resulted from the scrubber operating with commercial lime. The first 6 mixes in actuality froze during the winter months and the remaining impoundments will have to go through that phase this winter. This effect will not be known until some time around the middle of 1978. Generally, Pool No. 1, though difficult to measure due to winter conditions, attained a permeability in the order of 10^{-4} at the start and decreased to the 10^{-5} and 10^{-6} range this summer. All other carbide lime impoundments at zero days were in the order of 10^{-5} and have decreased to 10^{-6} as time has elapsed. Pool No. 3 and Pit No. 2 seem to be approaching what we would classify as satisfactory limits.

In the case of the commercial lime sludge mixes, the mixes as placed in the impoundments were of a higher percent solids and thus at zero days had permeability values in the order of 10^{-6}. Present indications, but not an actual condition due to the short time since placement, appear to be decreasing to the 10^{-7} area. All of the disposal impoundments resulted in mixtures with unconfined compressive strength in excess of 0.5 ton per square foot. Mix No. 11 and 12, which are similar in type to Mix No. 2 in Pool No. 3 and Pit No. 2, have relatively high unconfined compressive strength values in the order of 4 tons per square foot, and may even be too hard for ready handling as land fill. Hardness in itself is not the only criteria in an overall evaluation of fixation.

Limited leachate data at the time of this paper is restricted to carbide lime mixes. The following data from the three lysimeters on the basically unlined Pit No. 2 appears to be following a typical trend at 180 days with permeability values in the order of 10^{-6}.

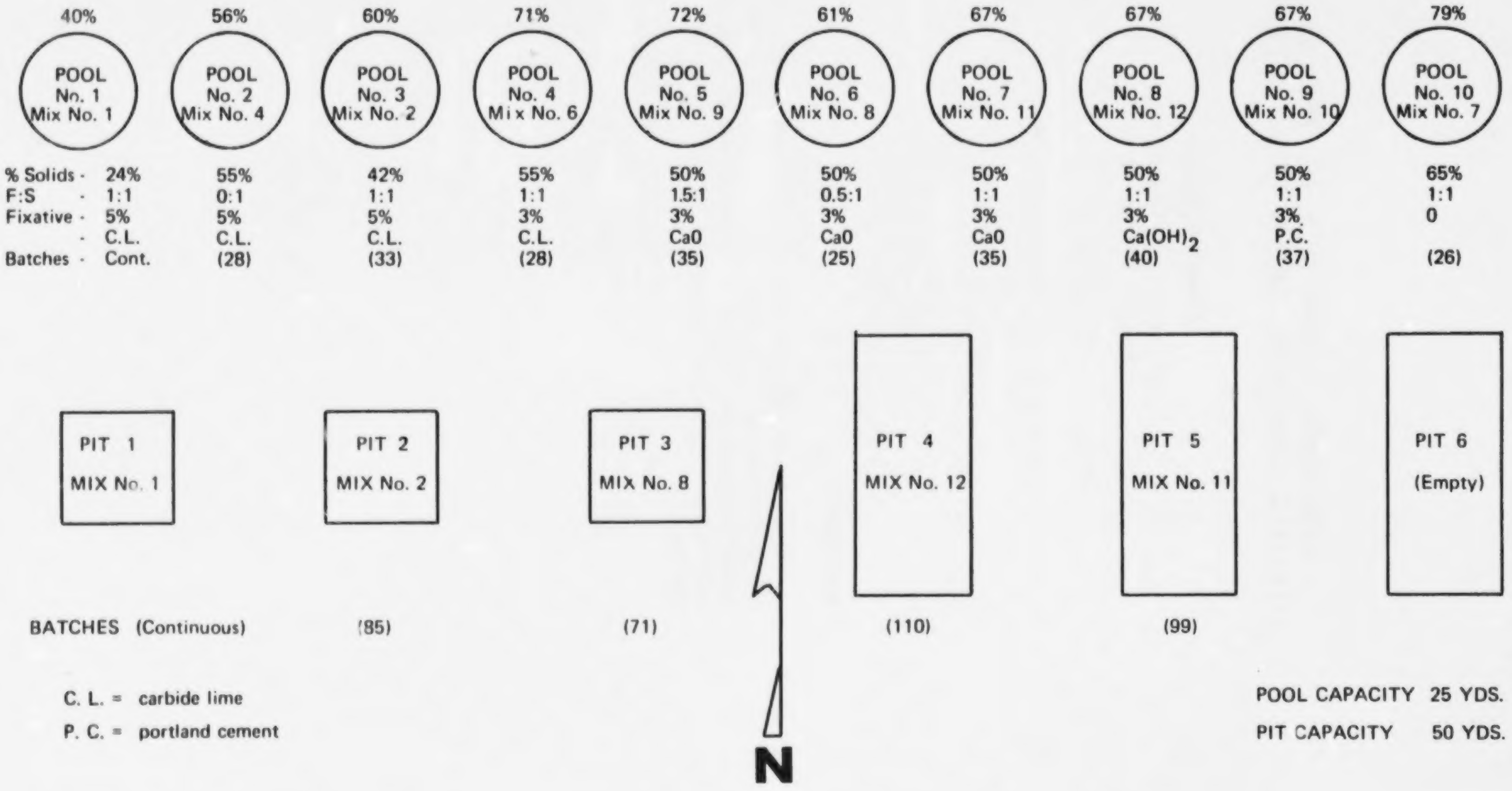
	Lysimeter 6" Below Mix Mg/L	Lysimeter 2' Below Mix Mg/L	Lysimeter 6' Below Mix Mg/L
180 Day Na	210	100	20
" " Ca	580	390	25
" " Mg	100	50	30
" " As	.006	.004	Below Detection Point
" " Cl	550	380	25

Real meaningful leachate data is too limited at this time to evaluate. In actuality, the entire evaluation of all 15 impoundments will not be realistically known until next summer when all the data will be available. For the moment, one can only say the carbide lime mixtures look good as do the lime mixes.

EXHIBIT A

SLUDGE IMPOUNDMENT SITE

IN PLACE ACTUAL PERCENT MIX SOLIDS



SCRUBBER EXPERIENCE AT THE KENTUCKY UTILITIES COMPANY GREEN RIVER POWER STATION

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ABSTRACT

Operating experiences with the American Air Filter lime scrubbing flue gas desulfurization system, which consists of an adjustable throat venturi scrubber and flooded elbow for particulate removal, and a compartmented mobile bed contactor for sulfur dioxide removal, are discussed.

SCRUBBER EXPERIENCE AT THE KENTUCKY UTILITIES COMPANY
GREEN RIVER POWER STATION

The Kentucky Utilities Company (KU) purchased from the American Air Filter Co., Inc. (AAF) a turnkey sulfur dioxide and particulate collection system in June, 1973. Startup commenced in September, 1975. The unit is a single module (Figure 1) sized to scrub the flue gas from three boilers which are headered to serve two 32 MW turbines. The scrubber module is designed (Table 1) to handle a maximum of 360,000 ACFM of flue gas at 300°F and contains an adjustable throat flooded approach venturi for fly ash removal, a mobile bed contactor (Figure 2) for SO_2 removal and a centrifugal demister. Pebble lime storage, slaking facilities and pumping system provide reagent slurry which is pumped to the three-compartment reactant tank. Bleed slurry 10-12% solids from the tank is pumped to a settling pond for permanent disposal and clear water from the pond is returned to the system for closed loop operation.

During the last months of 1975 and early 1976 the scrubber was operated at reduced loads. The turbines were out at different times for overhaul and maintenance. The scrubber was undergoing a shakedown and adjustment. The bypass was utilized more than it should have been. If it wasn't the turbines, or the scrubber it was one of the boilers that had a problem.

In April 1976 the decision was made to not operate the boilers without the scrubber except in an emergency (Table 2 and Table 3). In December 1976 an attempt to perform an efficiency test on the scrubber failed. It was found that the boilers had excessive air leakage and couldn't be operated at capacity.

In February, 1977 the scrubber was shut down to repair the stack where the Carboline lining had failed and to repair the boilers. In June 1977 the plant operators went on strike and since then there has been no operation.

In the two years since start up both AAF and KU have corrected and solved a number of operating and mechanical problems.

This scrubber is no different than others when it comes to scaling. It has experienced plugged spray nozzles, plugged mobile beds, scale buildup downstream of the demister and in the demister. There usually are reasons for these problems which AAF and KU are endeavoring to find the answers.

When the units return to service the scaling problem will be addressed by cycling the mobile bed dampers to prevent stagnation zones in the unit and the removal of the spray nozzles to increase the liquid flow to the unit and to prevent the piping from plugging with solids.

Acid fallout from the plume necessitated paying compensation to a number of employees for damage to their cars. This fallout also attacked the superstructure of the nearby substation. To avoid further damage, KU has authorized AAF to design and install a stack gas heating system. This proposal will utilize extractive steam from an adjacent unit to heat air and inject it into the stack to raise the exhaust temperature some 50°F.

The Carboline lining in the stack failed around nearly one half of the circumference. One half of the stack for its entire height was covered with 3/8 inch steel plate. The stack was then lined with 3/4 inches of Precrete over a wire mesh.

AAF is trying to improve the performance of the present centrifugal demister. If this cannot be done, it will be replaced with a Chevron design demister.

Other small problems which have been solved are: replacing the smaller balls in the mobil bed contactor with larger balls to reduce ball migration, replacing the rubber covered pump impellers with a Ni-hard impeller, installing straightening vanes to reduce stack vibration caused by the centrifugal demister, and redesigning the degritter to reduce grit pick up.

There are operational problems that require an operators attention and know how. On shut down the recycle pump screens have to be blanked off to prevent their being plugged. Instruments require attention or they will drift out of calibration. The guillotine dampers are difficult to close in zero weather. The automatic slaker and line system can plug up without attention.

As in any type of scrubber manpower is most important. The experience that is gained by the operators as time goes by will provide a more efficient operation.

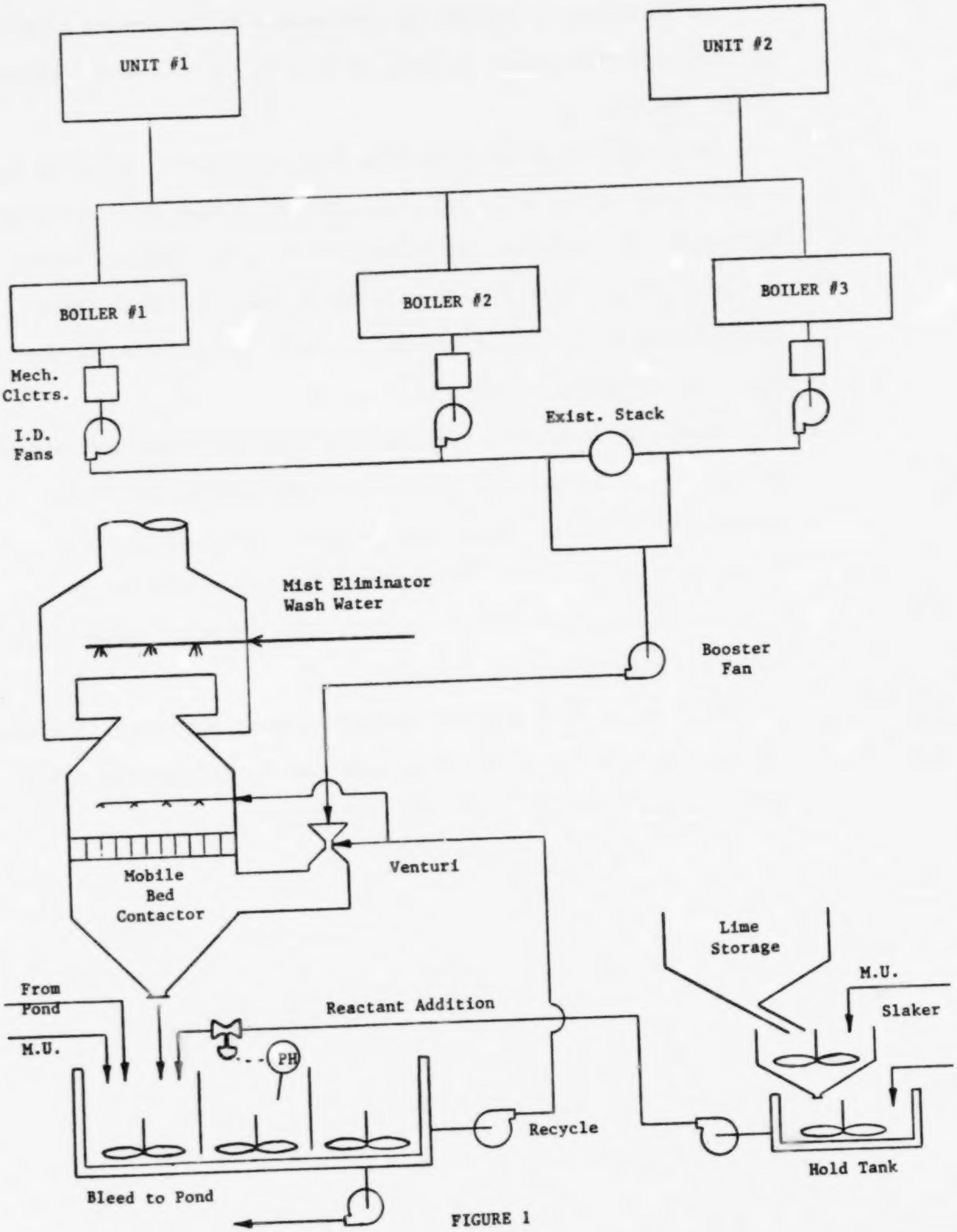
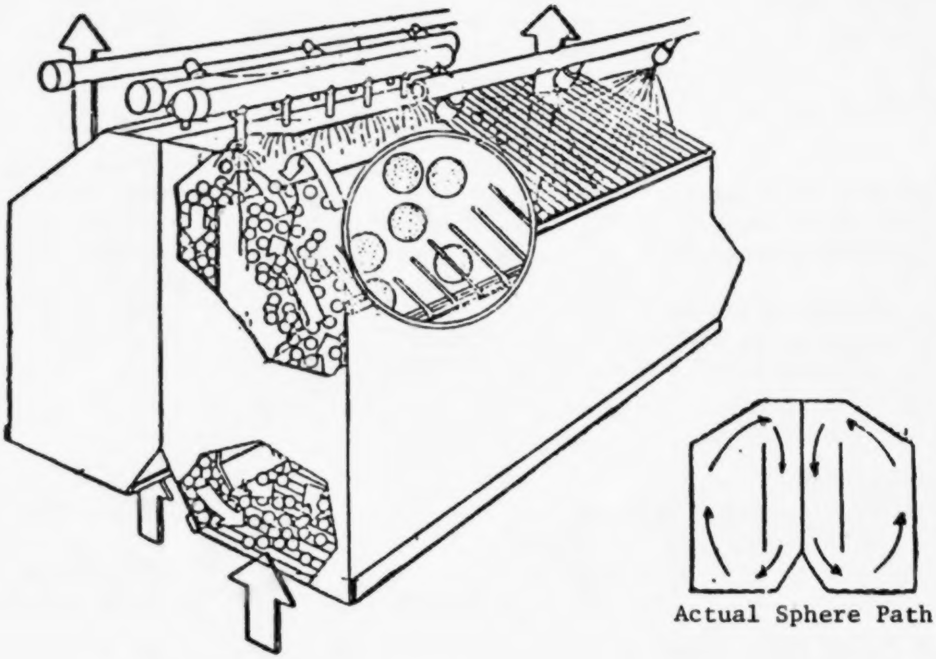

UNIT #1
UNIT #2
BOILER #1
BOILER #2
BOILER #3
Mech.
Clctrs.
I.D.
Fans
Exist. Stack
Mist Eliminator
Wash Water
Booster
Fan
Venturi
Mobile
Bed
Contactor
Lime
Storage
M.U.
Slaker
From
Pond
M.U.
Reactant Addition
PH
Recycle
Hold Tank
Bleed to Pond

FIGURE 1



Mobile Bed Contactor

Actual Sphere Path

FIGURE 2

TABLE 1

Kentucky Utilities Company - Green River Power Station
American Air Filter Designed Particulate and Sulfur Dioxide Flue Gas Scrubber

Two 30 MW (net) generating units (60 MW) served by three 215,000 lb/hr steam B&W-FH boilers 910°-900PSIG burning 3.98 lb coal/1000 ACFM of flue gas.

Average Coal Analysis

Moisture	10.0%	Sulfur-3.8%
Ash	14.0%	BTU - 11,000

Gas Stream Data

	L/G=gal/1000 cuft	39.5
Prior to Fan	SCFMDG	238,000
	ACFM @ 300°F, 14.69 PSIA	360,000
	Particulate loading/SCFDG	2.2gr
	SO_2 lb/min	108.9
In Final Stack	SCFMDG	238,000
	ACFM(sat.) @ 116°F	303,000
	Particulate loading/SCFDG	0.044 gr=1.5 lb/min
Particulate Removal efficiency		98%
Particulate Emission Level Priority I-lbs/MMBTU		0.145
Particulate Emission Level at full load-lbs/MMBTU		0.097
	SO_2,lb/min.	21.8
SO_2 removal efficiency		80%
SO_2 emission level Priority II-lb/MMBTU		2.0
SO_2 emission level at full load-lb/MMBTU		1.67

Liquid and Solids Data

To Contactor		
From Reaction Tank	H_2O	11,800 GPM
	Fly Ash	--------
	$Ca(OH)_2$	109 lb/min.
	Ca_xSO_x (approx.)	7,380 lb/min.
From Sump, GPM		20 GPM
Sprays from Pond		75 GPM
Vapor in Gas Stream		73 GPM
Total H_2O in		11,968 GPM
Leaving Contactor		
To reaction tank - GPM H_2O		11,800 GPM
In stack gas - GPM H_2O Vapor		168 GPM
Total H_2O Out (GPM)		11,968 GPM
To Pond	H_2O	228 GPM
	$Ca(OH)_2$ (maximum)	9.0 lb/min.
	$CaSO_x$	190 lb/min.

Utilities - Approximate Operating Requirements

Fan - One	1500 BHP
Recycle Pumps - Two	500 BHP
Mixers - Three	150 BHP
Bleed Pumps - One	5 BHP
Miscellaneous Other	28 BHP
TOTAL Peak Operating Power	2183 BHP

TABLE 2

GREEN RIVER POWER STATION
OPERATIONAL DATA FGD UNIT

1976

MONTH	A Hours in Period	B Hours FGD System Available	C Hours FGD called upon	D Hours FGD System Operated	E Hours Boilers Operated	B/A % Avail-ability	D/C % Respond-ability	D/E % Oper-ability	D/A % Utiliza-tion	Unit % Load Factor
Jan.	744	312.00	456.00	64.00	571.55	41.9	14.0	11.2	8.6	55.2
Feb.	696	486.17	499.38	210.75	499.38	69.9	42.2	42.2	30.3	40.7
March	744	721.72	408.66	386.38	457.53	97.0	94.5	84.4	51.9	43.7
April	720	648.00	552.00	552.00	552.00	90.0	100.0	100.0	76.7	50.2
May	744	606.18	455.88	455.88	455.88	81.4	100.0	100.0	61.2	44.1
June	720	720.00	596.43	588.85	596.43	100.0	98.7	98.7	81.8	62.3
July	744	665.85	583.53	574.43	583.53	89.5	98.4	98.4	77.2	51.2
August	744	722.45	744.00	722.45	744.00	97.1	97.1	97.1	97.1	54.0
September	720	617.20	571.20	571.20	571.20	85.7	100.0	100.0	79.3	32.5
October	744	744.00	698.55	698.55	698.55	100.0	100.0	100.0	93.9	37.7
November	720	720.00	704.25	704.25	704.25	100.0	100.0	100.0	97.8	51.4
December	744	539.31	591.48	517.20	535.52	72.5	87.4	96.6	69.5	46.5
Year	8784	7502.88	6861.36	6045.94	6969.82	85.4	88.1	86.7	68.8	47.5

TABLE 3

GREEN RIVER POWER STATION
OPERATIONAL DATA FGD UNIT

1977	A Hours in Period	B Hours FGD system Available	C Hours FGD called upon	D Hours FGD System Operated	E Hours Boilers Operated	B/A % Avail-ability	D/C % Respond-ability	D/E % Oper-ability	D/A % Utiliza-tion	Unit % Load Factor
January	744	698.29	744.00	698.26	744.00	93.9	93.9	93.9	93.9	56.5
February	672	242.80	266.12	242.80	266.12	36.1	91.2	91.2	36.1	32.8
March	744	0	0	0	0	0	0	0	0	0
April	720	288.00	166.82	164.00	166.82	40.0	98.3	98.2	22.8	9.4
May	744	735.65	526.55	513.27	526.55	98.9	97.5	97.5	69.0	34.4
June	720	720.00	34.38	34.38	34.38	100.0	100.0	100.0	4.8	1.3
July	744	744.00	0	0	0	100.0	0	0	0	0
August	744	744.00	0	0	0	100.0	0	0	0	0
September	740	740.00	0	0	0	100.0	0	0	0	0
October	744									
November										
December										
Year										

CONVERSION OF THE LAWRENCE NO. 4 FGD SYSTEM

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and

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ABSTRACT

The limestone injection scrubbing system installed in 1968 on unit No. 4 at Kansas Power and Light Lawrence power station was recently replaced with a sulfur oxide control system that involves use of limestone in a spray tower absorber. The rationale for changing the system, the basis for the design, and the significant design features are described. Startup activities are discussed and results of initial operation are presented. System performance has not been fully evaluated.

INTRODUCTION

At the last Flue Gas Desulfurization Symposium held in New Orleans in March 1976, Mr. Daric Miller of The Kansas Power and Light Company reported on the recent scrubber experience at the Lawrence Energy Center. Since the meeting in New Orleans, the Lawrence No. 4 scrubber system conversion has been completed and the scrubber has gone into service. This paper will describe the system, relate the operating experience to date, and report the results of recent performance tests.

SYSTEM DESCRIPTION

The Lawrence Energy Center Unit No. 4 is rated at 125,000 KW and burns Medicine Bow coal from Wyoming, which has a heating value of 10,000 Btu/lb, a sulfur content of .5 percent, and an ash content of 9.8 percent (Table I).

The air quality control system (AQCS), redesigned and installed by Combustion Engineering, is shown in Figs. 1 and 2. This system, comprised of two 50 percent capacity scrubber modules, is the first application of the C-E - RS/ST (Rod Scrubber/Spray Tower) two-stage AQCS. Table II summarizes the predicted performance.

The flue gas leaving the existing air heater is conveyed via new ductwork to the inlet of the rod scrubber. After passing through the rod scrubber, where the fly ash is collected, and the spray tower, where the major portion of sulfur dioxide is removed, the clean flue gas is demisted and reheated. It is then conveyed to the induced draft fan and pushed up the stack. Bypass ducts are provided, one for each module, so that the scrubbers can be bypassed when natural gas or oil is being fired in the boiler. Isolation dampers are located at the inlet and outlet of each module as well as in the bypass ducts. Flue gas flow is controlled by the I.D. fans.

Rod Scrubber

The rod scrubber (Fig. 3) is comprised of a converging section and a rod section. The converging section directs the downward flowing flue gas to the rod section, which has two staggered levels of rubber coated fiberglass rods. The rod section is 3 feet wide by 23 feet long. The rods each have a nominal diameter of 7 inches and are located on 13-inch centers. The vertical spacing between the two rows of rods is varied automatically in proportion to gas flow (load) in order to maintain a set gas side pressure drop across the rods. The rod section is sprayed with slurry continuously by non-atomizing fan type nozzles located around the perimeter of the rod scrubber. After the slurry passes through the rod section, it falls to the rod scrubber collection tank directly below. The slurry is recycled from the collection tank to the rod scrubber.

Experience at the Sherburne Co. AQCS of Northern States Power had led C-E to conclude that a sacrificial wear plate was required directly below the nozzles in the rod scrubber. These plates of 316L SS were manufactured but not installed prior to startup because of lack of time. They were installed during the April outage. It is anticipated that these wear plates will give approximately two years life.

Spray Tower

The flue gas makes a 180 degree turn after leaving the rod scrubber and enters the spray tower. The gas flows upward through the open tower and is contacted by slurry which is sprayed counter-current to the gas through two levels of sprays. The first spray level (Fig. 4) is ten feet above the inlet duct, the second spray level is ten feet above the first for a total contact zone of twenty feet in which the sulfur dioxide is removed. The spent slurry falls into the reaction tank where fresh limestone is introduced into the system. The slurry is recycled to the spray tower after it is held in the reaction tank a sufficient length of time (10 minutes) to enable precipitation of calcium salts and dissolution of fresh additive.

Mist Eliminator

The entrained moisture in the clean flue gas is removed in the mist eliminator section (Fig. 5). The mist eliminator is comprised of a bulk entrainment separator (BES) followed by a two-stage chevron demister. The mist eliminator is made of reinforced fiberglass which is stable at a temperature of 400 F.

Reheater

The flue gas is then reheated by an in-line carbon steel reheater. The reheater design incorporates a circumferential finned staggered tube arrangement. The reheater is four rows deep. This design is the same design which has operated satisfactorily for over five years on the existing Lawrence No. 5 AQCS.

Effluent Bleed

This system was designed with separate slurry hold tanks for the rod scrubber and spray tower. Staging of the liquid side of the system enables addition of fresh limestone to the absorber (spray tower reaction tank). The solids in the reaction tank are controlled at 5%. Slurry is bled from the reaction tank to the collection tank. The collection tank percent solids are controlled at 8 to 10% by varying the effluent bleed pump flow. This pump delivers the effluent bleed to the system thickener where the slurry is concentrated to 30 to 35% solids before being pumped to a disposal pond. Water from the pond is returned to the recirculation tank where it is combined with the thickener overflow water to provide makeup water for the scrubber system.

Additive System

The additive for the scrubber system is prepared on-site using a wet ball mill (Fig. 2). Three-fourths inch (gravel size) limestone is stored in a hopper. From the hopper the limestone is fed by a feeder to a wet ball mill where it is ground to an 80 percent through 200 mesh slurry. The 60% slurry from the mill is pumped to the additive storage tank. The additive is then transferred to the additive dilution tank via a variable speed pump. The additive is diluted with makeup water to maintain 10% solids in the additive dilution tank. The limestone additive is fed to the spray tower reaction tanks at a rate proportional to the unit firing rate and sulfur content of the fuel.

In-Tank Strainer

Each reaction tank and collection tank is supplied with an in-tank strainer. The strainer (Fig. 6) prevents oversized particles from entering the spray system and potentially plugging spray nozzles. Each strainer is equipped with an automatic water washer which back washes the strainer and prevents it from becoming plugged. These oversized particles are purged from the system via the effluent bleed which is located upstream of the strainer.

Cleaning Devices

Sootblowers are located at the inlet to the rod scrubber to prevent buildup at the wet/dry interface. Compressed air at 200 psi is the blowing medium.

Water washers are used with the in-tank strainer as mentioned previously as well as at the mist eliminator section. High pressure (80 to 100 psig) water is sprayed periodically (once per day) and automatically on the BES and demister. Figure 7 shows the intensity of this high pressure, high flow washer.

Additionally, two half-track sootblowers are located upstream of the reheater. Using 200 psig air as the blowing medium, they maintain the reheaters in a clean condition.

Materials of Construction

The scrubber is manufactured of 316L SS with a moly content of 2.7% from the rod scrubber inlet to the reheater. The ductwork downstream of the reheater is carbon steel. The slurry hold tanks are unlined carbon steel. The spray pumps are completely rubberlined, and the spray piping is reinforced fiberglass (FRP). The internal spray headers are also FRP.

Instrumentation

The C-E - RS/ST was supplied with a significant amount of instrumentation beyond what is normally required for routine operation. For example, magnetic flow metering devices were installed in all slurry lines, sulfur dioxide probes were located between the rod scrubber and spray tower, and slurry control valves were located in each of the spray lines of the RS/ST.

SYSTEM OPERATION

The system was put into service on January 23, 1977. It operated almost continuously until mid April when the unit was taken out of service for a scheduled outage. The unit was returned to service in May and the scrubber operated until June when the unit began firing gas and the scrubber system was bypassed. The unit began firing coal again in late September. To date, the scrubber system has logged over 3164 hours, since its startup in January.

SYSTEM PERFORMANCE

In April 1977 C-E conducted some preliminary tests to determine the performance of the AQCS in removing particulate and sulfur dioxide. These tests verified that the system was performing satisfactorily, but they also revealed a bypass damper leakage problem (discussed later). It was determined that a complete system performance test would be performed in October 1977 to determine the particulate and sulfur dioxide removal and opacity of the system. In addition, evaluation of the spray tower and rod scrubber at different operating conditions was recommended by C-E. Table III gives the results of these tests.

In general, these tests have established that the C-E - RS/ST can meet better than 99% particulate removal at 10 inches ΔP across the rods and can in fact produce an emission rate about 20% below 0.1 lb/mm Btu when operating under optimum conditions.

The opacity measurements were found to be between 2 and 8% for the twin eight-foot diameter stacks. These readings are extremely low, but appear to be consistent with the small amount of fine material in the outlet dust loadings and the stack diameter.

The SO_2 removal varied from 75 to 90+% depending on liquid to gas ratios and additive flowrates. This performance agrees with C-E's design program predictions.

Also, chemical analyses of the process streams were taken to establish a material balance as well as determine additive use in the AQCS. Table IV gives a summary of these analyses.

Figures 8 and 9 are plots of the particulate emission as a function of ΔP and opacity.

Detailed interpretation of this test data has not been completed at this time. It is anticipated that a paper covering analysis of this data will be issued jointly by KPL and Combustion Engineering at the spring 1978 meeting of Missouri Valley Electric Association.

OPERATING EXPERIENCE

Initial operation of a new system reveals the need for some field adjustments and/or design changes. Some of these items and the action taken are described below.

A. Freezing - The scrubber system is entirely outdoors and there was some freezing during initial operation in mid-January. It was not until the end of February that heat tracing and insulation was completed. Subsequently, it was decided to build enclosures around the spray pumps to prevent freezing when out of service and to allow for better maintenance in future winters.

B. Spray Pumps Seals - Several of the spray pumps required repacking every two or three days. Discussions with the pump vendor have resulted in redesign of the gland seal arrangement. The major change involved the incorporation of high flow gland seal water (14 to 15 gpm as opposed to the 7 to 8 gpm originally used).

C. In-Tank Strainer - The washing device for the strainer failed several times and required overhauling. It was determined that most of these failures were due to mechanical malfunction of limit switches, mal-operation, or human error. In addition, the cavity behind the in-tank strainer became plugged during periods when the system was shut down. This problem was caused by an inoperable agitation system.

The agitation is provided by forcing compressed air into the cavity (or spray pump suction well). The compressor is now operational. It is expected that this has resolved the problem.

D. Spray Pipe Failure - A failure in the FRP was discovered in May. It was determined that the failure was directly downstream of a butterfly valve that was throttling flow to the spray tower sprays. It was decided to open the valve completely, thereby eliminating the turbulence and resultant wear. The spray pressure increased from 18 to 23 psi which did not cause any additional problems with the process.

E. Mist Eliminator Wash Pump - It was determined after a week or two of operation that there was insufficient water pressure to the mist eliminator washers. This low pressure resulted in the washers being operated twice every 24 hours to ensure demister cleanliness. This additional wash water made it difficult to maintain the desired % solids in the reaction tank. Several times during the first two months of operation periods of scaling occurred. The scaling was limited to a total of 1/8 inch buildup in the absorber.

A higher pressure water source was hooked up temporarily until a booster pump could be installed during this past summer.

F. Damper Leakage - During some preliminary tests in April, it was determined that the bypass dampers did not seal completely, thereby allowing some particulate matter to bypass the AQCS. The bypass damper is a guillotine on one module and a compound louver on the other.

New seals were installed to minimize this leakage. Further, it has been determined that the damper drives drift. This means that every 5 to 10 minutes the dampers move off their limit switches. The controls are activated and the damper is driven back to its correct position. This problem allows flue gas to bypass the scrubber and causes nuisance alarms in the control room. A redesign of the damper drive mechanism is being supplied by the damper manufacturer.

G. Rod Scrubber Inlet Blower - Initial operation of the system indicated that the inlet sootblower was not cleaning the wet/dry interface adequately. The lance was modified to obtain better coverage and has functioned properly since that time.

H. Additive Pump Wear - The limestone additive is pumped to the module via a positive displacement screw type pump. The pumps provide an excellent means of controlling the additive feedrate. In fact, on this particular unit which requires 8 to 32 gpm of 10% limestone slurry, they appear to be the most acceptable means of controlling additive. Operation at this installation as well as C-E experience at other sites indicates that the freshly ground limestone is highly abrasive. The pump liners and rotor wear out in 10 to 15 weeks of service. This life expectancy is short, but since there are only two of these pumps in the system and the wear can be predicted, it has been decided to operate with the existing additive pumps rather than evaluate the cost effectiveness of another type of additive feed control.

I. Mixers - Several of the rubber coated blades of the top entry mixer in the collection tank experienced stabilizer fin failure in February. They were replaced by the manufacturer of the mixers and no additional failures have occurred. There were several bearing failures of the side entry mixers in the reaction tank. These failures have been traced to improper lubrication.

Most, if not all of the above problems have been resolved. It is significant to note that no cleaning of the scrubber modules has been required due to plugged nozzles or scaling. Obviously, when a mechanical component such as a mixer or washer fails to operate, not only does that piece of equipment have to be repaired, but it produced some additional maintenance during the time it is out of service.

The overall operation of the AQCS to date has been very encouraging. The process problems that plagued the original installation have been eliminated, and the selection of materials of construction in the new system gives every indication of being satisfactory for power plant installation.

This information is significant because the same design philosophy has been incorporated in the Jeffrey Energy Center Unit 1 (720 Mw) AQCS, which will commence operation in early 1978, and the conversion of the Lawrence Energy Center Unit 5 (400 Mw) AQCS, which will be completed in April 1978.

CONCLUSIONS

The startup and initial operation of the converted Lawrence 4 flue gas desulfurization system has been satisfactorily completed. Most problems have been resolved and performance of the system meets prior predictions.

Though the new system has eliminated most of the problems associated with the original Lawrence 4 AQCS, it is still a major piece of operating equipment, and therefore will require attention while operating as well as an increase in the plant's maintenance requirements.

This AQCS, which uses the RS/ST (Rod Scrubber/Spray Tower) for particulate and sulfur dioxide removal, is a major improvement in flue gas desulfurization system design. The system affords essentially infinite turndown capability, few scrubber internals, and high performance capabilities.

TABLE I

MEDICINE BOW COAL

ULTIMATE ANALYSIS	TYPICAL
Moisture	11.8
Carbon	60.7
Chlorine	.03
Sulfur	0.55
Ash	9.8
Btu (As rec'd)	10,000
ASH ANALYSIS	
Silica	38.0
Ferric oxide	9.5
Alumina	23.9
Lime	13.2
Magnesia	3.5

TABLE II

PREDICTED PERFORMANCE OF LAWRENCE 4
AIR QUALITY CONTROL SYSTEM

(Medicine Bow Coal)
(0.9% Sulfur, 9.8% Ash)

Superheater outlet flow, lb/hr	842,100
Kilowatts	125,000
Gas entering particulate scrubbers, lb/hr	1,290,000
Gas entering particulate scrubbers, CFM @ 280 F	403,000
Gas entering I.D. fans (2), lb/hr	1,339,000
Gas entering I.D. fans (2), CFM @ 144 F	363,000
Excess air leaving economizer, %	25
Gas reheater - Rise in Temp, F	20

DRAFT LOSSES, inches wg

Boiler and air preheater	8.6
Ducts	2.6
Particulate scrubber	9.0
Absorber	2.0
Mist eliminator	0.5
Gas reheater	0.5
Stack	0.8
TOTAL	24.0

LIQUID FLOW RATES

Additive feed total, lb/hr	3000
Additive feed total, GPM @ 10% Solids	56
Spraywater flow per module	
Particulate scrubber, GPM (L/G=20)	3600
Absorber, GPM (L/G=30)	5300
Effluent bleed per module	
Particulate scrubber, lb/hr	6696
Particulate scrubber, GPM	125
Absorber, lb/hr	2110
Absorber, GPM	40
Solids Concentration, % by weight	
Particulate scrubber	9-11
Absorber scrubber	5-7
Makeup water total (Fresh Water), GPM	130
SO_2 concentration particulate scrubber, inlet, PPM	748
SO_2 concentration absorber outlet, PPM	200
SO_2 removal, %	73
Particulate removal, %	98.9

TABLE III

PERFORMANCE TEST RESULTS

Date	10/18/77	10/10/77	10/12/77	10/12/77	10/14/77	10/18/77	10/18/77	10/19/77	10/20/77	10/20/77	10/24/77	10/24/77	10/25/77	10/25/77
Test No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Location	South outlet	South outlet	South outlet	South outlet	South outlet	South outlet	South outlet	South outlet	South outlet	South outlet	South outlet	South outlet	South inlet	South inlet
Particulate loading, GR/SCF D	.031	.022	.030	.024	.026	.028	.031	.032	.035	.032	.039	.039	3.077	2.990
Particulate loading corrected to 3% O_2, GR/SCF D	.044	.031	.043	.034	.037	.040	.044	.046	.050	.046	.056	.056	4.39	4.27
Opacity, %	2.5	3.0	2.5	3.0	2.5	2.5	2.0	2.0	2.0	2.0	7.5	7.5	--	--
Rod section ΔP, inches H_2O	10.4	10.1	16.0	16.0	10.4	10.4	10.4	10.4	16.0	16.0	4.5	4.6	4.5	4.5
L/G RS/ST	20/30	20/30	20/30	20/30	10/30	10/30	20/0	15/0	20/0	20/0	20/30	20/30	20/30	30/30
Excess air, %	64.7	67.6	63.3	63.3	61.5	64.8	60.2	60.2	61.9	63.9	68.4	68.4	68.4	68.4
Gas temp, F	145	143	142	144	146	145	144	144	144	144	147	147	288	292
Gas flow, CFM	228,254	229,301	236,000	231,948	238,554	228,275	224,444	228,951	231,845	235,691	138,475	137,363	153,575	156,623
Load, Mw	51	52	52	52	53	51	51	51	51	51	51	51	52	52

TABLE IV

CHEMICAL DATA

TEST	A	B	C
Limestone Feed, % Stoichiometry	100	41	18
Inlet SO_2 (1)		(400 - 450 ppm)	
Outlet SO_2 (1)		(10 - 20 ppm)	
% SO_2 Removal		(90+)	
% Oxidation - CT/RT	78(2)/98(2)	58/95	57/98
% Limestone use - CT/RT	60/38	75/67	81/87
% Solids - CT/RT	8.5/7.8	12.4/8.0	12.8/5.2
pH - CT/RT	7.5/6.6	6.8/6.3	7.7/5.5
Chemical Analysis (3)			
Liquid			
Ca (ppm)	876/715	801/702	781/669
Mg (ppm)	137/127	225/210	256/214
SO_3 (ppm)	106/23	100/87	89/214
SO_4 (ppm)	2340/2064	2570/2375	2598/2303
$CaSO_4$ Relative Saturation	1.45/1.22	1.38/1.21	1.35/1.15
Solids			
$CaSO_3 \cdot \frac{1}{2}H_2O$ (% wt)	2.41/.20	6.50/1.35	6.85/0.38
$CaSO_4 \cdot 2H_2O$ (% wt)	11.57/19.25	11.65/28.70	11.78/30.65
$CaCO_3$ (% wt)	5.85/21.52	3.74/8.57	2.83/2.35

(1) Corrected to 3% O_2

(2) During Test A air was added to both tanks to increase oxidation. The amount of air was equivalent to approximately 600% stoichiometry.

(3) All chemical analyses are reported for the rod scrubber collection tank/spray tower reaction tank.

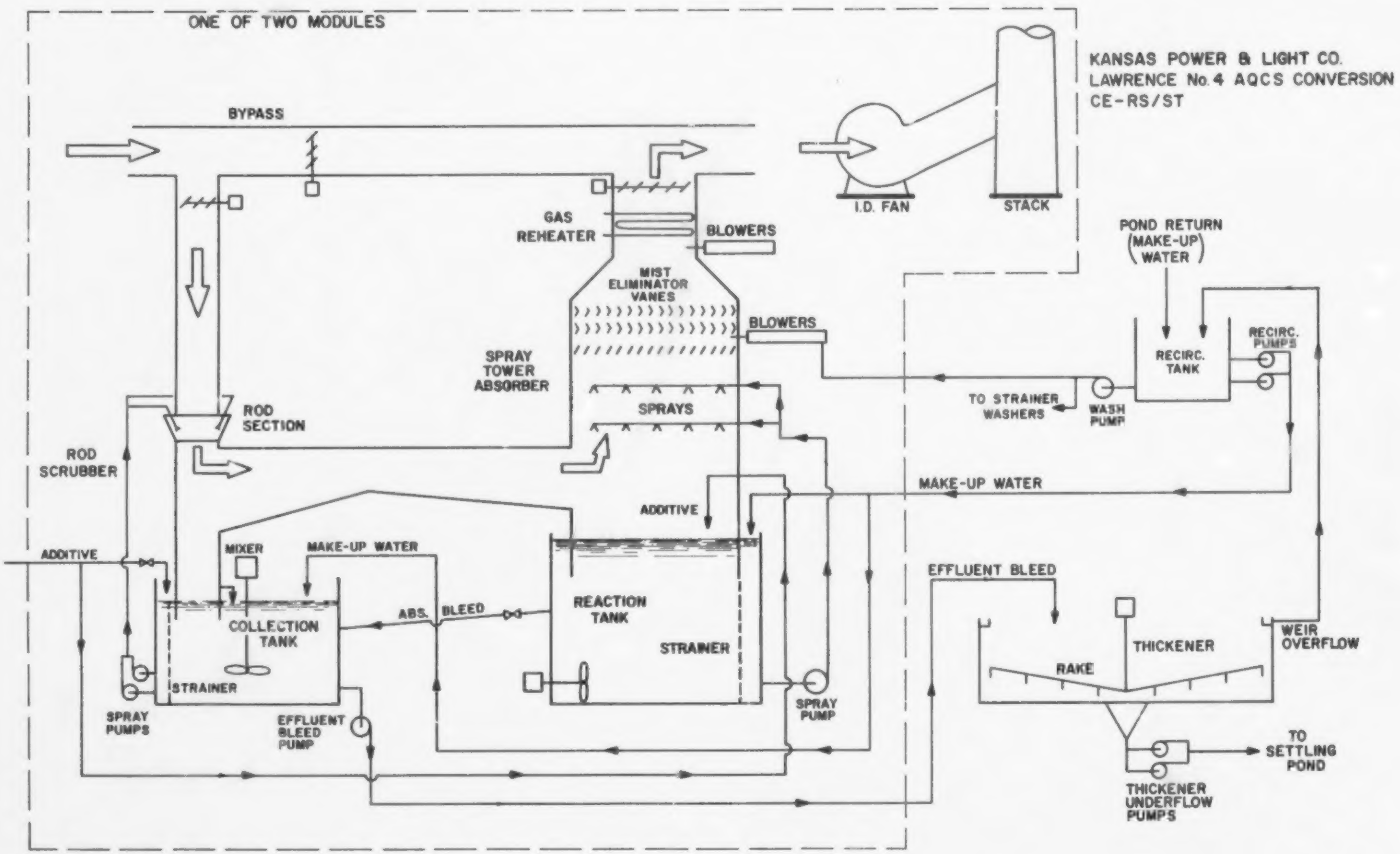

Figure 1: Kansas Power and Light Company Lawrence No. 4 AQCS conversion C-E - RS/ST

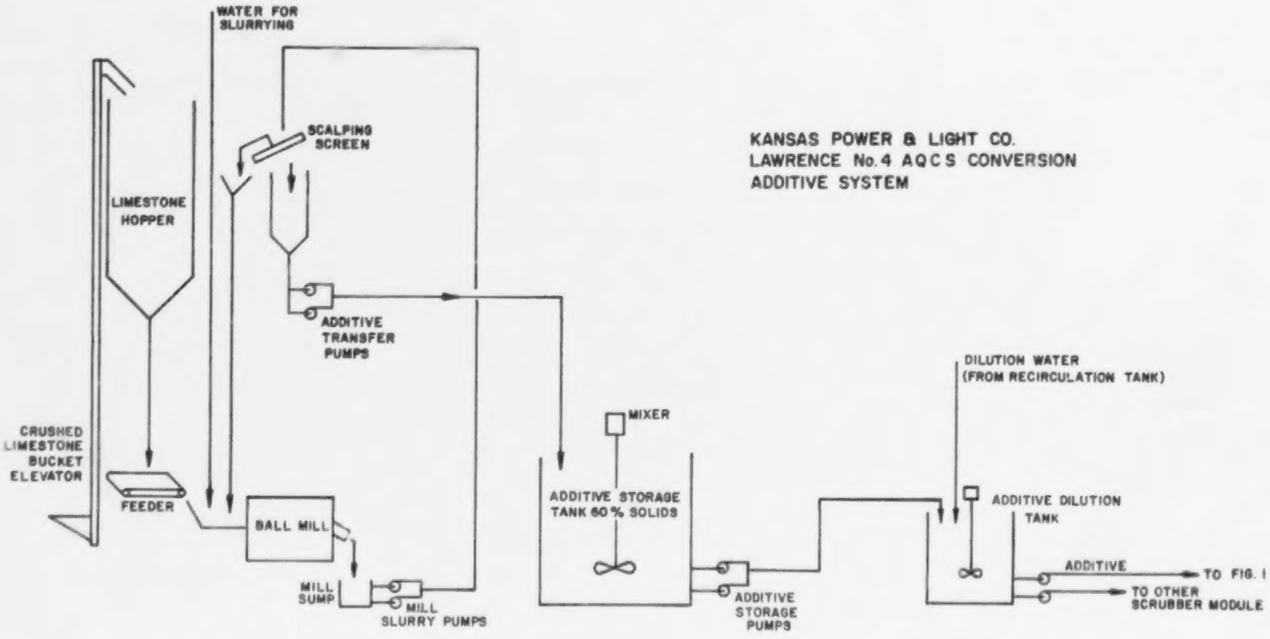

Figure 2: Kansas Power and Light Company Lawrence No. 4 AQCS conversion additive system

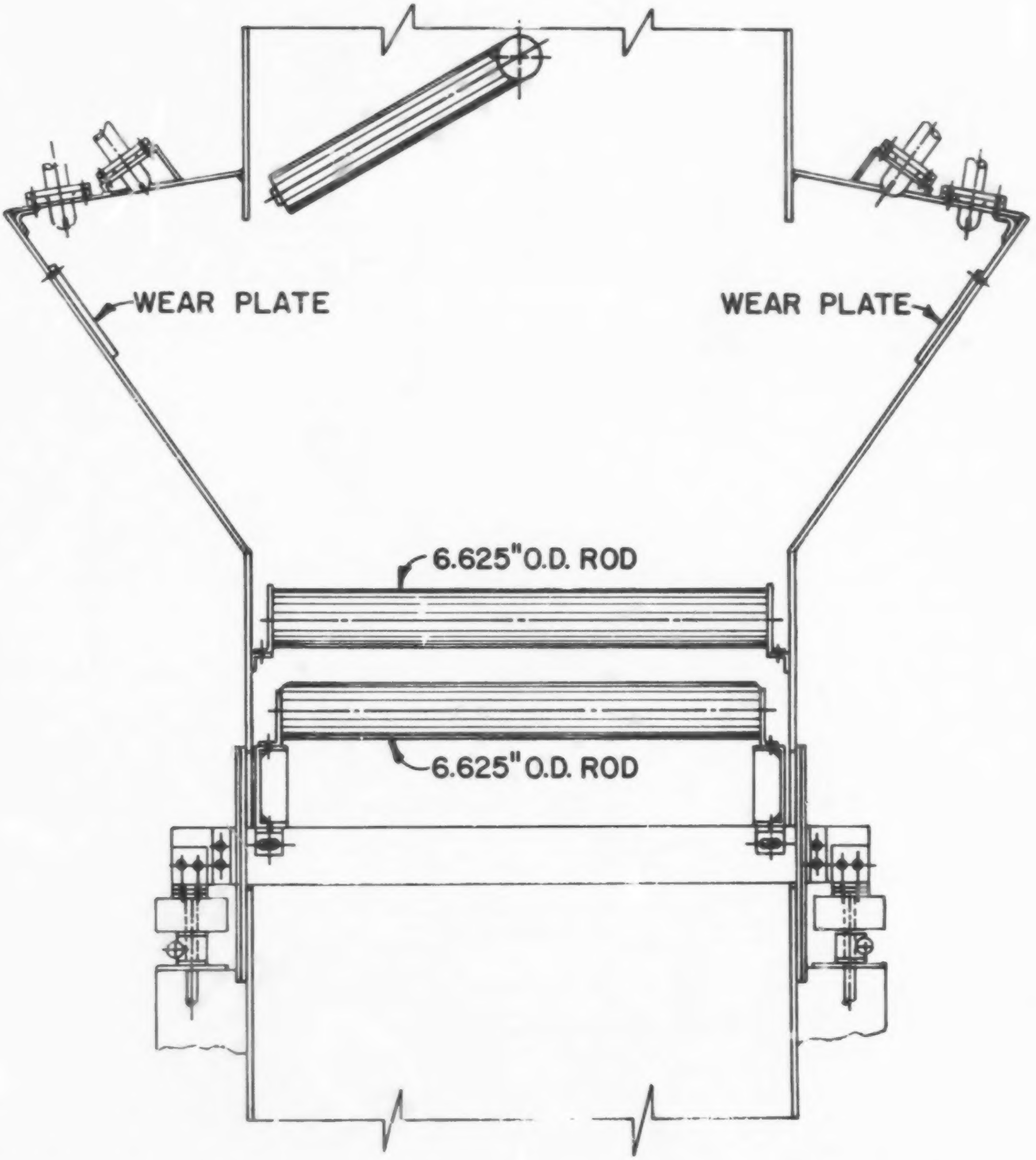

Figure 3: Variable throat rod section

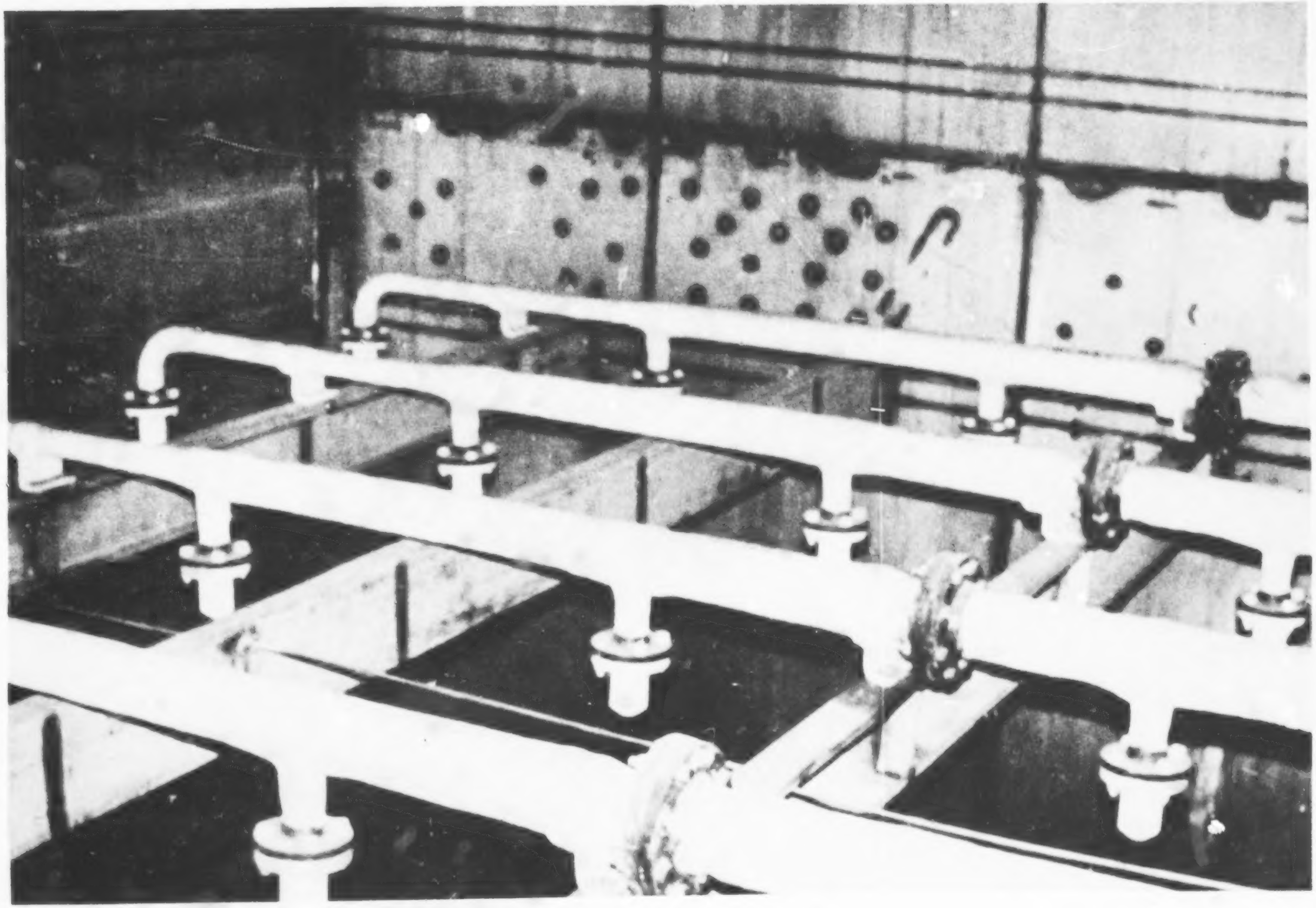

Figure 4: First spray level in spray tower

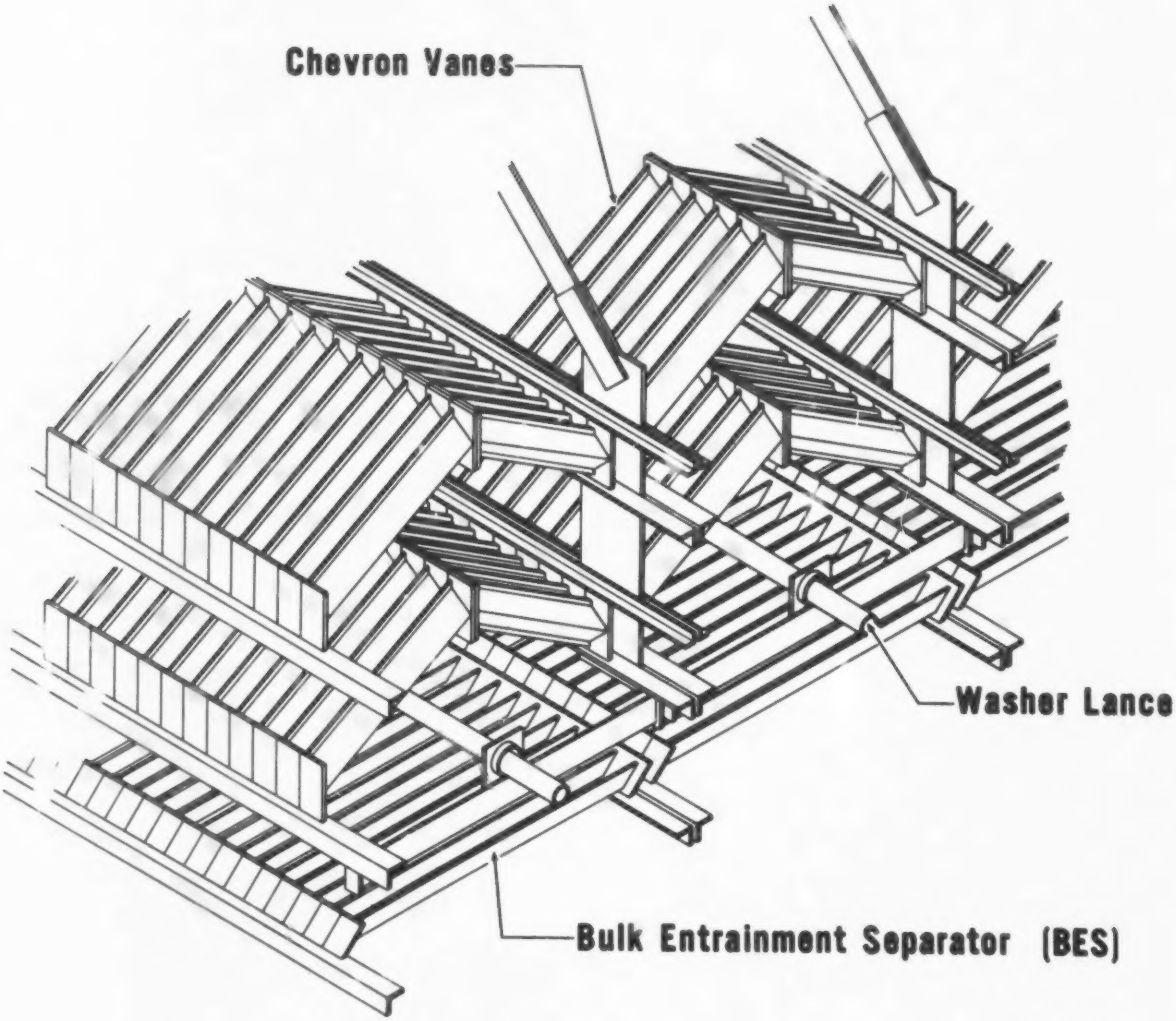

Figure 5: Mist eliminator system

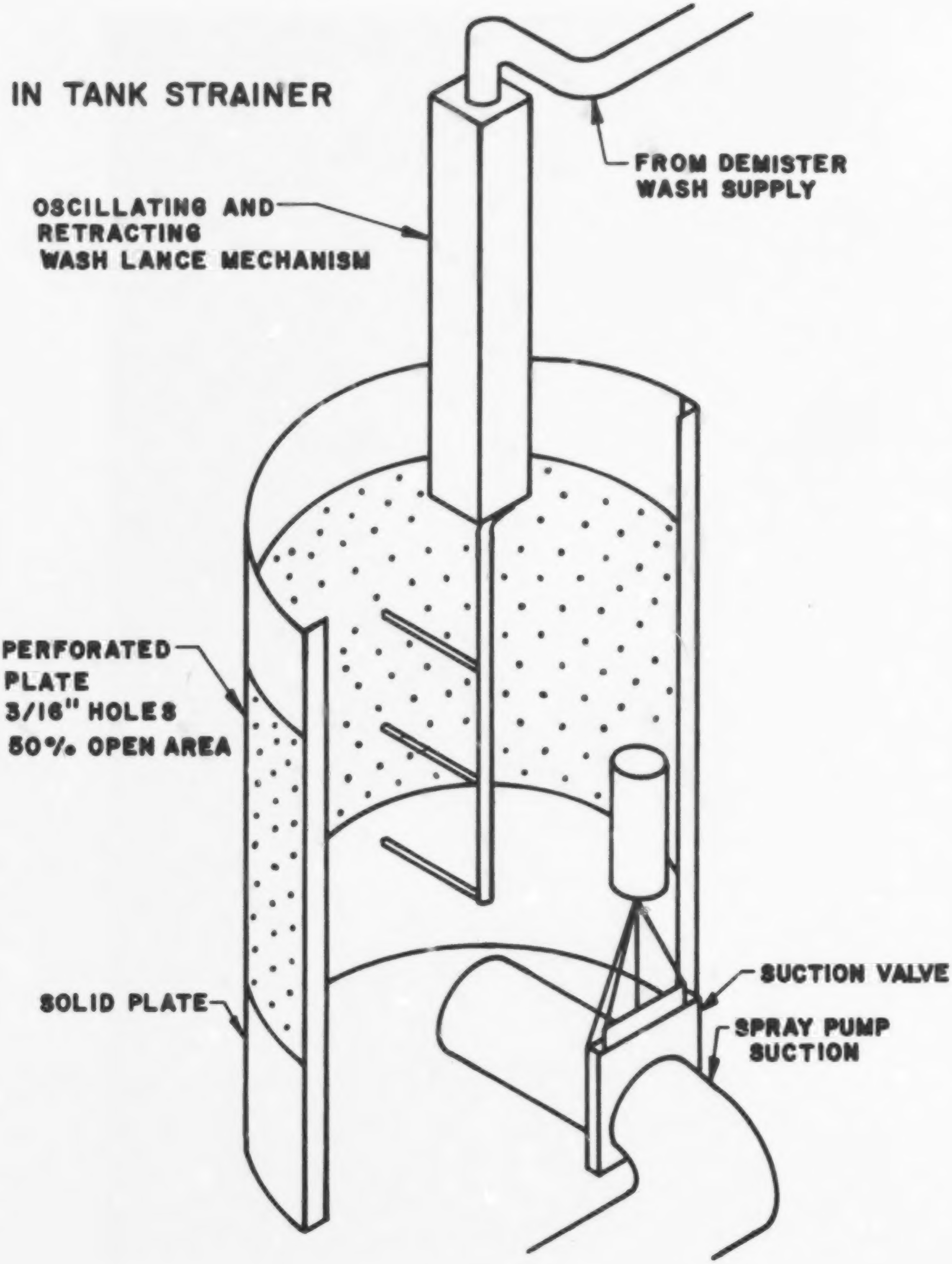
IN TANK STRAINER
FROM DEMISTER WASH SUPPLY
OSCILLATING AND RETRACTING WASH LANCE MECHANISM
PERFORATED PLATE 3/16" HOLES 50% OPEN AREA
SOLID PLATE
SUCTION VALVE
SPRAY PUMP SUCTION

Figure 6: In tank strainer

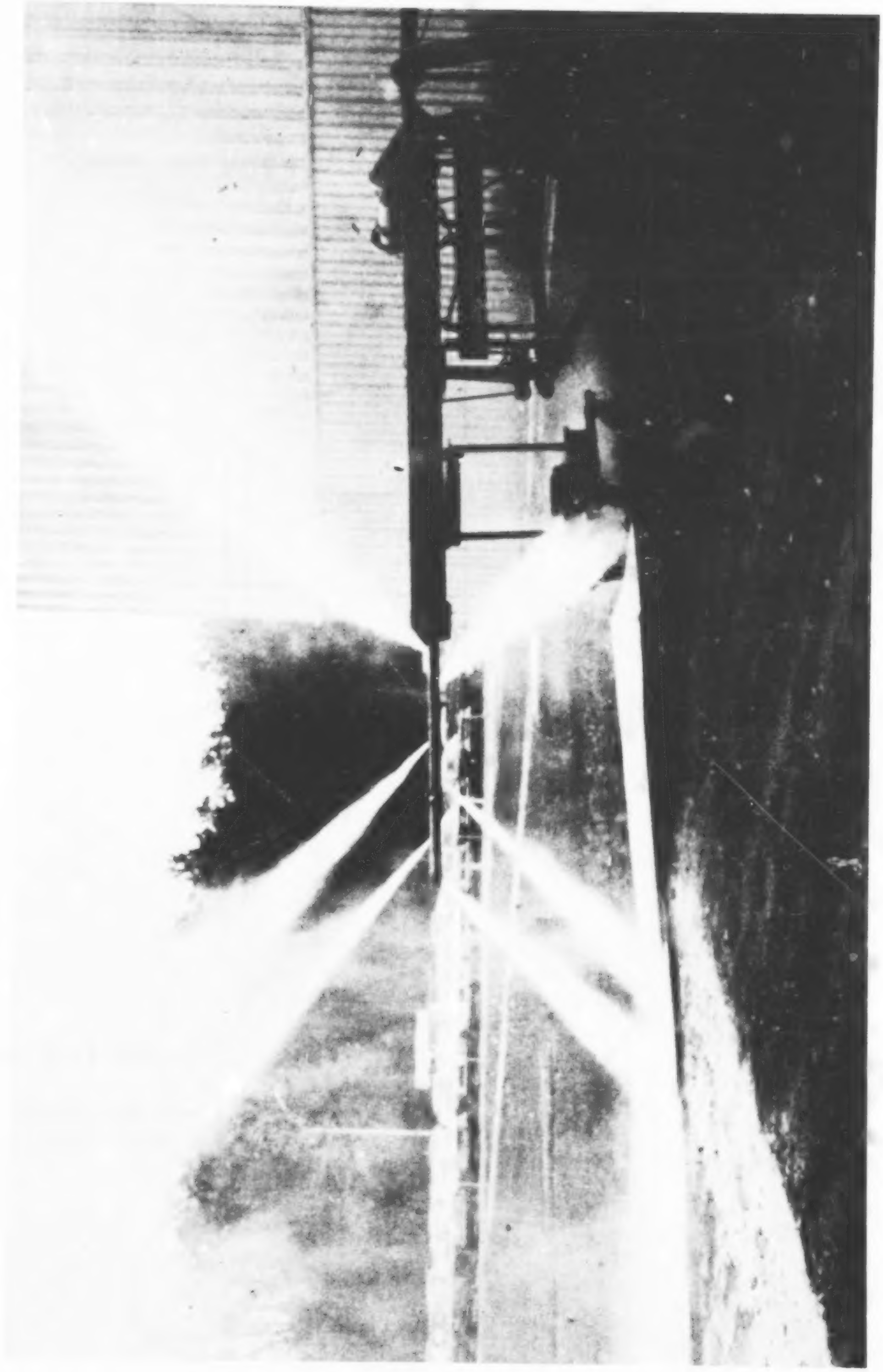

Figure 7: High pressure water washer

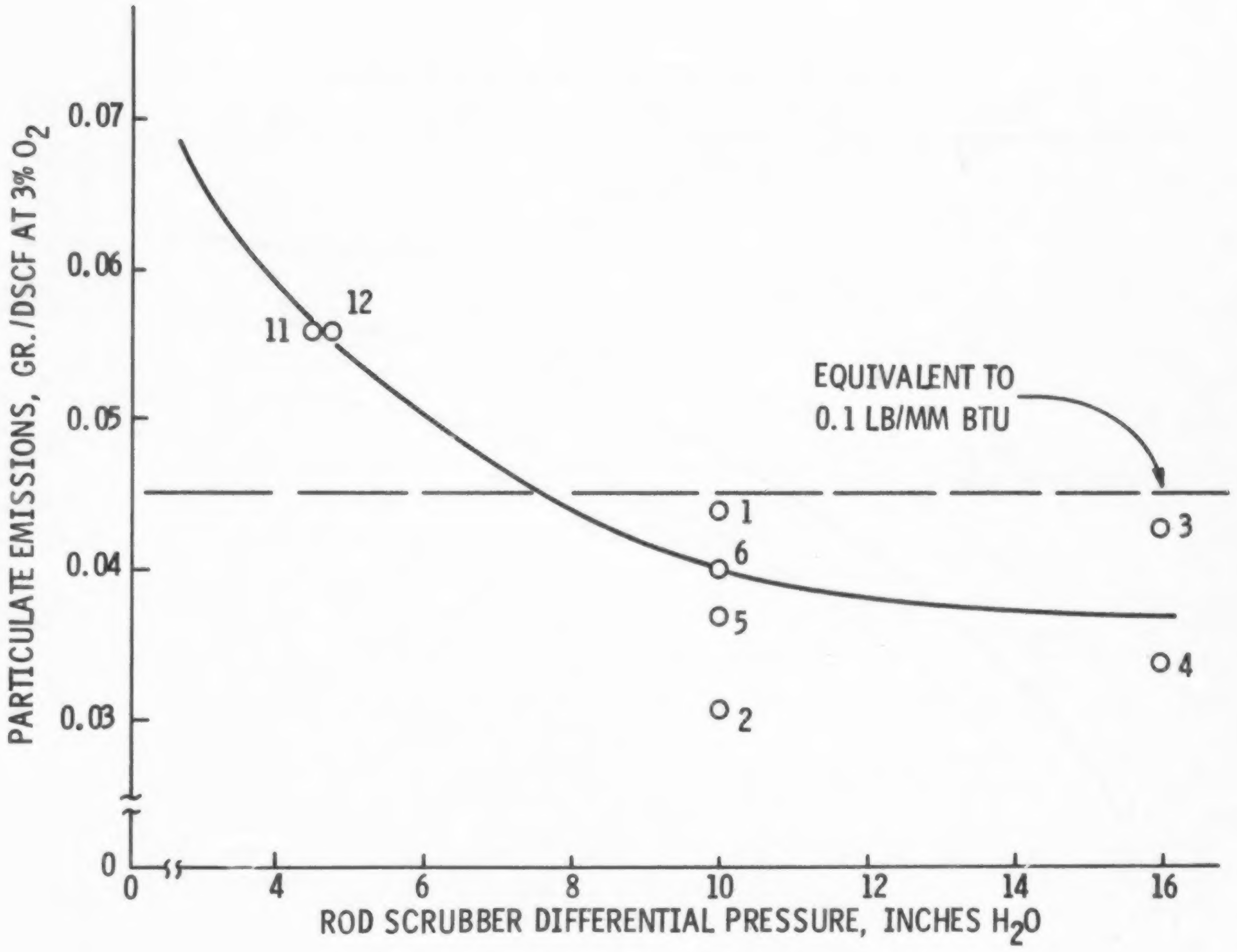

Figure 8: Particulate emission vs rod scrubber differential pressure

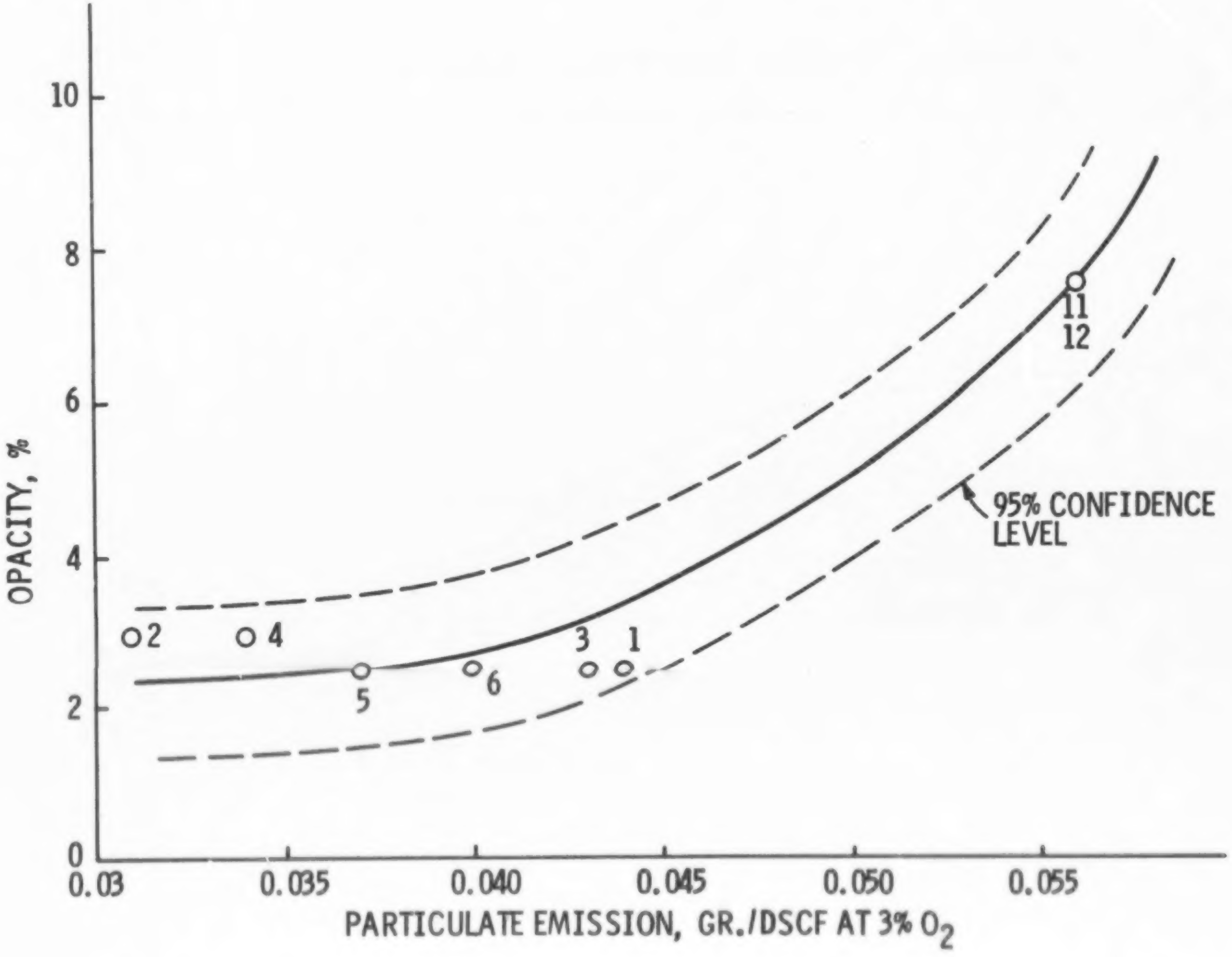

Figure 9: Particulate emission vs opacity (for 8 ft. diameter stack)

STATUS AND PERFORMANCE OF THE MONTANA POWER COMPANY'S FLUE GAS DESULFURIZATION SYSTEM

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ABSTRACT

Sulfur dioxide and particulate removal systems on the two new 360 MW coal-fired units owned jointly by The Montana Power Company and Puget Sound Power and Light Company have successfully demonstrated performance within the Federal New Source Performance Standards. Unit #1 has been in operation since September 1975 and Unit #2 since May 1976. Three scrubber modules, each containing a variable throat venturi for particulate and sulfur dioxide removal and a spray absorption section for additional sulfur dioxide removal, serve each generating unit. No bypass capability is installed. Each 33-⅓ percent nominal scrubber module is capable of treating up to 40 percent of the flue gas flow for short periods of time.

This system, which was piloted over a 2-year period, uses the alkalinity present in the Rosebud seam coal for the major portion of sulfur dioxide removal. Supplemental lime is added for pH control. The system design is based upon a maximum of 1 percent sulfur in the coal.

The scrubber system components, design parameters, and the operational performance are discussed in the paper.

STATUS AND PERFORMANCE OF THE MONTANA POWER COMPANY'S FLUE GAS DESULFURIZATION SYSTEM

INTRODUCTION

The two mine-mouth 360 MW coal-fired generating units located at Colstrip, Montana included in their initial design combined flue gas desulfurization and particulate removal equipment. The scrubber systems on both units were built with the intention of meeting the Federal New Source Performance emission standards. Due to the timing of the new standards, Unit #1 appeared to be exempt from them. However, the State agency responsible for issuing the permit for the Colstrip project expressed its desire for both SO_2 removal and particulate control so both units were designed and built to meet the NSPS.

The generating units are located next to the coal fields approximately 30 miles south of the Yellowstone River in Southeastern Montana. The subbituminous coal is taken from the Rosebud seam. The ash of the coal possesses significant alkalinity and the design intent for the scrubber system was to make use of this characteristic.

The FGD system employed at Colstrip operates in the closed loop mode. Liquid can only leave the system by evaporation in the process up through the stack, evaporation from the pond surfaces, or be purged with the sludge resulting from the scrubber operation.

Successfully incorporating the FGD system into the design of the full scale plant meant that a pilot study had to be run for an appreciable time to obtain design parameters and provide reliability data. Pilot operations on the 3,000 acfm scale were conducted over approximately two years at the J. E. Corette Plant of The Montana Power Company in Billings, Montana. Colstrip coal was burned in the furnace similar in design but smaller than the ones that were to be constructed at Colstrip. Operation of the pilot proved the initial premise that alkalinity present in the ash could be used to react with the sulfur dioxide present in the flue gas; the pilot investigation also showed that integration of a wash tray into the design was required to successfully keep the mist eliminators clean, yet live within the water balance required for closed loop scrubber operation. Operator training on the pilot was also a side benefit from the program and these operators performed duties on the full scale units when they went into operation.

The vendor for the scrubber installation was Combustion Equipment Associates. Bechtel Power Corporation was the architect-engineer for the plants and personnel from its Research Division were heavily involved in the scrubber developed for Colstrip.

DESCRIPTION OF SYSTEM

Each of the generating units is served by three scrubber trains. There is no bypass capability. Two of the trains can handle 80% of the combustion gases for a limited time period should maintenance be required on one of the trains.

The design parameters of the FGD system are as follows:

1. Liquid to gas ratio: L/G 33
   split L/G 15 to venturi, L/G 18 to absorption sprays.
2. Suspended Solids in Recycle Loop: 12%.
3. Venturi Pressure Drop: 17.0 in. W.G.
4. Recycle Retention Time: 8 minutes.
5. Recycle Slurry pH: 4.5 to 5.6.
6. Particulate Emission Guarantee: Outlet grain loading not to exceed 0.018 grains per actual cubic foot as measured at the reheater outlet.
7. Sulfur Dioxide Guarantee: SO_2 emission not to exceed 1.0 lb SO_2 per million Btu heat input.

The emission guarantees were deliberately chosen as stated above because they could be directly related to the standards which are written as emission limit in pounds of emissions per million Btu, not efficiencies.

Table 1 describes the coal and ash characteristics as used in the scrubber specifications.

Figure 1 is a schematic representation of the scrubber train. Combustion gas exits the air heater at about 290°F, travels through a duct containing several 90° turns and enters the scrubber module. Figure 2 presents a detailed view of the scrubber internals and Figure 3 supplements with a schematic view of the associated piping and pump arrangement. The gas then passes through the variable throat venturi where it is contacted with the slurry which flows over both the venturi bowl surface and the plumb bob. Next, it passes through the absorption zone where it is contacted with recycle liquor sprayed from the absorption nozzles. The wash tray is next in line to contact the gas. Its function is to remove entrained solids from the cleaned flue gas prior to the mist eliminators. Both the bottom of the mist eliminator and the bottom of the wash tray are sprayed continuously. The mist eliminator is sprayed with a blend of wash tray pond return liquid and river water.

TABLE 1

FUEL AND ASH AS DIRECTED IN SPECIFICATIONS

	Average As Received
Coal:	
Moisture	23.87%
Volatile Matter	28.59%
Fixed Carbon	38.96%
Ash	8.59% (Max 12.58%, Min 6.1%)
Heating Value	8,843 Btu/lb (Min 8,162 Btu/lb)
Sulfur	.77% (Max 1.0%, Min 0.4%)
Ash: (Estimated composition, sulfur trioxide-free basis.)	
SiO_2	41.60%
Al_2O_3	22.42%
TiO_2	0.79%
Fe_2O_3	5.44%
CaO	21.90%
MgO	4.95%
Na_2O	0.31%
K_2O	0.13%
P_2O_5	0.41%

(Balance unidentified.)

Later data varies slightly from above.

Figure 1

COLSTRIP UNITS 1 & 2
SIMPLIFIED FLOW DIAGRAM

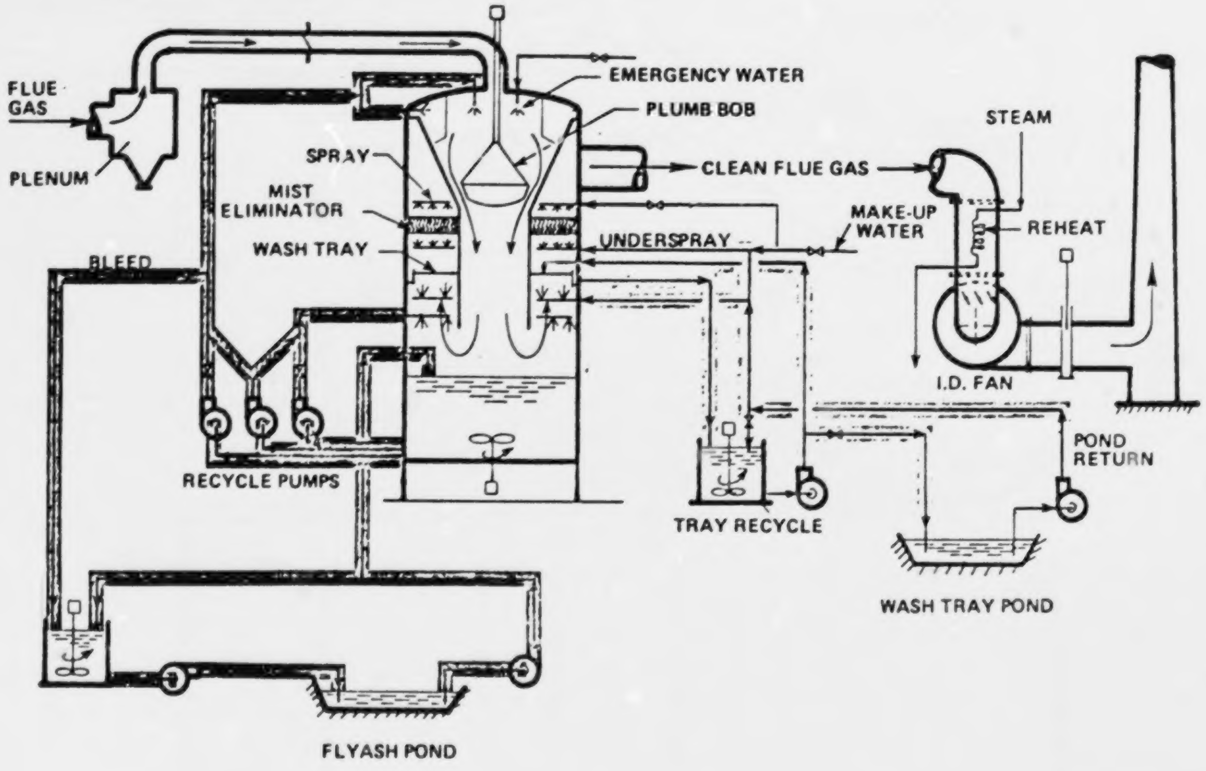


Figure 2

TYPICAL FGD SCRUBBER
COLSTRIP UNIT 1 & 2

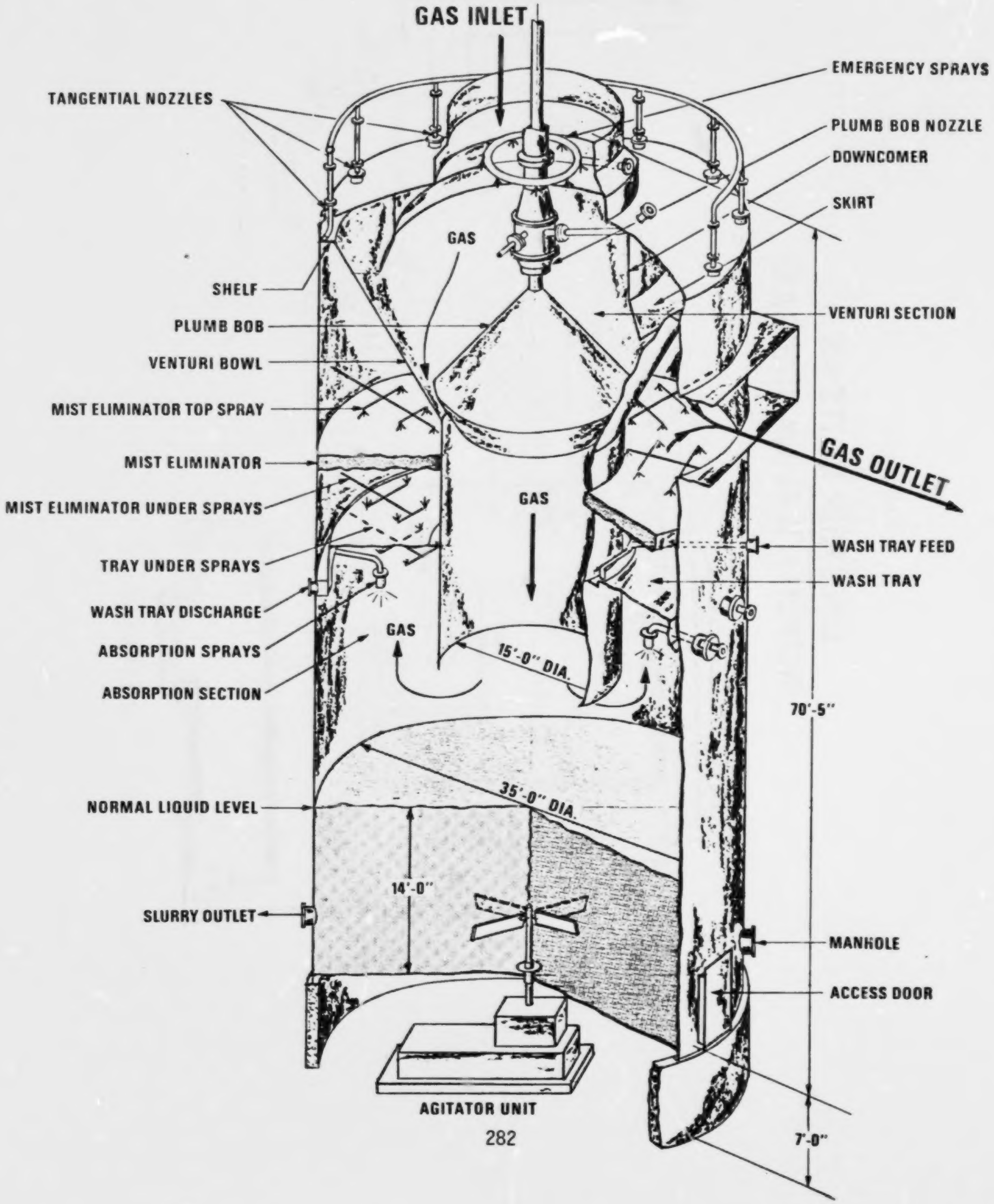
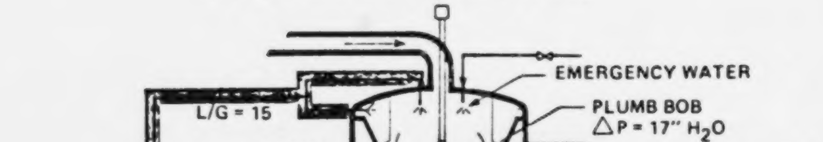


Figure 3

COLSTRIP UNITS 1 & 2
SCRUBBER MODULE



This area and the pump and agitator seals are the only places where fresh water enters the system. A liquid level is maintained on the top of the wash tray. The wash tray/mist eliminator loop works independently of the scrubber slurry recycle loop. The gas exits the scrubber module and passes through a reheat section consisting of two banks of plate-coil reheaters. The reheaters are cleaned by soot blowers placed above and between the two banks. The dry induced draft fans are last in line to contact the gas, then it passes isolation dampers and out the stack.

Two scrubber recycle pumps are needed to provide the required slurry flow to the venturi and absorption sections. The third pump is a spare that can be valved in to either the venturi or the absorption sprays. A bottom entry agitator in the main recycle tank provides necessary agitation to extract alkali values from the ash and resuspension capability for the slurry should the module be shut down. The wash tray tank is also agitated and a set of pumps, one operating and one spare, is used in that loop. The three scrubber modules of one generating plant rely on one tray recycle tank.

Solids are bled from the scrubber recycle loop to the effluent tank, then pumped to the settling ponds immediately behind the plant. The wash tray cycle rejects its solids to a separate pond at the same location. Decanted liquid is brought back to the wash tray and scrubber recycle loops.

Materials of construction include unlined carbon steel for the scrubber inlet ducts, glass flake polyester lined carbon steel for the internal surfaces of the scrubber vessels, abrasion resistant bricks in the venturi throat, 316 stainless steel on certain areas of the plumb bob, a Norel plastic for the four pass chevron mist eliminators, 316 stainless steel in the wash tray, flakelining in the duct exiting from the scrubber through to the chimney opening, Hastalloy G and Inconel 625 for the reheat surfaces with 316 stainless forming the duct walls around the reheater and rubber lining in the fan casing. The large recycle pumps are all rubber lined. Piping four inches and greater in diameter is rubber lined carbon steel and piping less than four inches is 316 stainless steel. The dry induced draft fan is of carbon steel construction.

The lime system consisting of storage silos for the pebble lime slakers, transfer and holding tanks, and the associated pumps and piping provides for lime slurry addition to the scrubber recycle tank for pH control.

Inlet and outlet sulfur dioxide monitors are provided on the scrubber trains. The testing station placed in the stack also is the location for the SO_2, NO_x, CO_2 and opacity monitors.

The scrubber installation is controlled from the plant main control room. Special attention during design was directed toward providing sufficient walkways, inspection ports and access hatches throughout the scrubber complex.

OPERATIONAL HISTORY

Unit #1 was first synchronized in September 1975 and declared in commercial operation during November 1975. Unit #2 followed with synchronization in May 1976 and commercial operation in August 1976.

The physical appearance of the plume from Unit #1 was good from the beginning of the operation, indicating that a high degree of particulate control was achieved. The dissolved solids built up rapidly in the associated scrubber ponds as the system operated in the closed loop mode. The scrubber system has operated in water balance with no surplus of water due to the scrubber operation. The scrubber has been free of the massive scale formations that have been reported in the literature on other installations.

The overall performance of the scrubber system has been quite good. There have been several problem areas, however. Instrumentation in slurry and flue gas service has not been trouble free. The stack opacity, SO_2 and NO_X monitors along with the SO_2 monitors on the inlet and outlet of the scrubber modules have exhibited unstable and erratic behavior. A corrective program has shown progressive improvement in the data retrieved by these monitors during the past six months. The in line pH probes either erode away or lose sensitivity due to deposits on the elements. Slurry density monitors have given erratic and unstable operation too and a test program is in progress to isolate these devices from vibration. Both slurry density and pH, key operating parameters, are taken manually and monitored on a frequent basis.

Deposits occurred at the wet-dry interface located at the entrance to the scrubber modules which reduced availability to an unacceptable degree. These are not to be confused with calcium sulfate or calcium sulfite scale formation. Gas flow tests were run on a model that duplicated the configuration of the duct from the air heater exit to the top of the scrubber module and the upper portion of the scrubber module itself. These tests found there was a maldistribution of the dust entering the scrubber module along with extreme velocity variation from one side of the duct exit to the other. In addition, a model study was made of the liquid flow on the tangential shelf above the venturi. From those studies, it was concluded that gas flow turning vanes in the duct elbow above the scrubber module and liquid guiding vanes and baffles on the tangential shelf would be required to reduce the wet-dry interface buildup. One module was modified and operation gave encouraging results. These modifications were then made during the past summer and spring to the remaining five modules and the construction time is reflected in the scrubber availability figures in Table 2. Actual operating experience from the modules equipped shows a significant reduction in cleaning time and has increased the time increment between inspection and cleanings.

The supplementary alkali feed system has been difficult to use due to line plugging and equipment failure. The lime system is being modified so that it will be simpler to use and less prone to plugging. Pinch valves are being placed next to the scrubber vessel on each feed line so any solid accumulation will slough off as these valves are regularly stroked. The alkaline ash captured by the scrubber has generally been adequate to keep the scrubber pH within the control range. Better control of the lime system will keep the pH more constant and allow SO_2 removal to be optimized.

Failures of the protective glass flakelining on the carbon steel vessel walls and ducts have occurred. These failures were evident prior to a major temperature excursion that occurred in Unit #1 on October 1976 but were then aggravated. The excursion followed a complete station power blackout and the failure of the emergency scrubber quench water supply system to operate. The glass flakelining along with the plastic mist eliminators were seriously damaged at that time in Unit #1 and account for low scrubber availability during November and December 1976. A failure of an ID fan motor during the same time period, independent of the temperature excursion, also contributed to the lower availability figures. If the ID fan motor was not considered in compiling the availability numbers, the scrubber availabilities would have been 88.5%, 94.0% and 81.6% for Unit #1 during the months of October, November and December 1976. Subsequent problems with other ID fan motors early in 1977 have reduced scrubber availabilities. The availability given for May 1977 for Unit #2 reflects another fan motor repair and the modifications for the wet-dry interface. Table 2 contains data on scrubber availability from September 1975 through July 1977. The definition of scrubber availability below the table should be noted.

Inspection of an ID fan rotor during the spring of 1977 resulted in the finding of cracks in the center plate next to the blades. Each fan was then cleaned and critically inspected. If cracks were found or if there was suspicion a crack might form, it was isolated and ground out. Then that portion was welded and a stiffener plate was added. Availabilities during the smmer of 1977 and the modifications carried out for the wet-dry interface show this activity.

Quick clean basket strainers were added to the suction piping of the two main recycle pumps of each vessel in place of the original start-up strainers contained in a pipe spool. This has allowed deposits and foreign materials to be removed before they clog spray nozzles and has also decreased maintenance and downtime on the vessels.

A program is under way to evaluate abrasion resistant protective lining materials that could be used where the flue gas makes its 180° turn following the venturi down-comer and passes by the absorption sprays. Impingement on the wall by the absorption spray slurry erodes the glass flakelining and this lining has been replaced at least once in each module since unit start-up.

TABLE 2

SCRUBBER AVAILABILITY VS PLANT LOAD

	Monthly Capacity Factor %		Number Days On Line		Average MW For Days On Line		Scrubber Availability %	
	Unit #1	Unit #2	Unit #1	Unit #2	Unit #1	Unit #2	Unit #1	Unit #2
Sep 1975	0.5		3		50			
Oct	19.4		19		139			
Nov	42.2		24		203			
Dec	59.9		30		239			
Jan 1976	63.8		28		265		90.0	
Feb	65.4		26		273		98.0	
Mar	57.0		24		277		97.6	
Apr	49.9		28		219		74.2	
May	26.0	1.3	14	3	210	66	96.8	100.0
Jun	0.0	23.2	0	16	0	171	-	99.7
Jul	28.0	19.5	20	13	167	180	93.2	98.7
Aug	37.8	13.0	23	10	194	162	94.7	95.8
Sep	64.5	64.6	30	30	239	232	88.6	98.3
Oct	73.1	77.0	30	31	281	298	79.9	90.3
Nov	55.6	79.7	30	30	225	303	62.7	94.1
Dec	67.2	82.3	31	31	249	297	73.8	92.5
Jan 1977	72.9	68.2	31	30	270	258	92.5	83.4
Feb	3.0	75.3	2	27	161	285	95.4	93.8
Mar	0	71.3	0	28	-	293	-	96.7
Apr	49.9	68.2	25	29	236	264	83.0	84.5
May	64.1	23.3	26	13	286	209	85.4	63.0
Jun	68.6	61.4	28	28	284	249	87.4	87.8
Jul	71.8	57.5	29	28	284	238	85.1	90.6

Note: Scrubber availability = total module hours available ÷ 3 (hours in month) except May through August 1976 base is days in operation because of extended scheduled outages.

Similarly, a program for evaluating corrosion, temperature resistant and abrasion resistant materials in other areas of the scrubber is in progress. Small test patches and full scale application of various linings are in place on several of the scrubber modules and are subject to different environments.

Plugging of the wash tray underspray nozzles on scrubber modules of both units occurred during May of 1977 and forced the plants to shut down. The major cause of the problem was traced to piping in the wash tray pond which had broken and allowed some of the pond dike material to be ingested. A parallel pipeline and pumping system was intalled, the units brought on line and modification and repair made to the original piping system.

The scrubber ponds are reclaimed by a dredge system which transfers the scrubber sludge to a disposal pond approximately three miles from the plant site. Liquid from the disposal pond is returned to the scrubber ponds immediately behind the plant when the dredge is not operating. The scrubber sludge deposits in deltas in the ponds and contain 55% - 60% solids. The material hardens very well and the dredge has had some difficulty in reslurrying the sludge. Residual alkalinity remains in the sludge deposited in these ponds and this results in the sludge taking on a cement-like nature. Improved cutters for the dredge and a delumper ahead of the dredge pump are expected to increase the solids handling capacity. More rapid filling of the wash tray pond with solids than had been expected has presented a problem in the method of removal of these materials. Each of the plant ponds has its own chemical identity due to water balance considerations. A dredge cannot be used presently to reclaim the wash tray pond volume without adversely affecting either the plant operation or the pond water balance, so a less efficient clamshell operation is now in use. Plans are being developed for the addition of another wash tray pond so that one may be dewatered then cleaned while the other is in operation.

TEST DATA AND EVALUATION OF SYSTEM

A large number of emission tests have been conducted on the Colstrip units over the 1-1/2 year period. FPA compliance tests for particulate, sulfur dioxide and NO_X were completed during the spring of 1977 on both plants. Also, emission monitor certification testing has been conducted following the conclusion of the NSPS emission tests. Table 4 shows the results of the tests. The NSPS requirements, the scrubber guarantee, the projected results from pilot plant experience and the actual tests on Units #1 and #2 are reported. The results of the operation of the full scale units agree well with the pilot plant data. The data also shows the plant emissions are well below the guarantee and the federal standards. Although scrubber inlet data has not been taken and efficiency per se is not available, it appears sulfur dioxide removal efficiency of 70% - 75% and particulate removal efficiency of 99.5% predicted from the pilot plant data are being achieved.

TABLE 4

EMISSION TEST RESULTS - EPA METHOD

	SO_2			Particulate			NO_x	
	LB/HR	PPM	LB/MMB	LB/HR	LB/MMB	% OPAC	LB/HR	LB/MMB
1. Required by NSPS (358 MW)	4063	510	1.2	339	0.10	20	2370	0.7
2. Scrubber Guarantee (358 MW)	3386	425	1.0	207	0.06	-	2370	0.7(1)
3. Projected from Pilot Plant (358 MW)								
a) 0.78% S (760 PPM), 8.19% Ash	1394	185	0.41	130	.038	20	2370	0.7
b) 1.0% S (965 PPM), 12.58% Ash	2071	260	0.61	184	.054	20	2370	0.7

4. Unit #1 Tests:

		Coal As Received(4)			SO_2			Particulate			NO_x	
Date	Gr MW	% Sul	% Ash	Btu/lb	LB/HR	PPM	LB/MMB	LB/HR	LB/MMB	% OPAC	LB/HR	LB/MMB
2/76	353	0.83	9.03	8638	1587	197	0.48	95.8	.029	10(3)	954	0.29
4/76	210(6)	0.71	7.79	8861	456	87	0.23	60.9	.030	12(3)	800	0.41
7/76	184(6)	0.64	8.49	8807	258	53	0.15	57.1	.031	15(2)	740	0.42
9/76	186(6)	0.62	7.93	8633	275	56	0.15	64.4	.037	11(2)	687	0.39
12/76	223(6)	0.94	8.54	8394	898	154	0.43	67.2	.032	15(2)	662	0.31
1/77	354	0.77	8.17	8719	770(5)	133(5)	0.35(5)	123.5	.037	13(2)	1459	0.43
5/77	331	0.61	8.41	8851	552	72	0.16	109.6	.032	14(2)	1029	0.29
6/77	340(7)	0.91	8.96	8888	1164	122	0.325	89.5	.039	18.4(2)	1296	0.360

5. Unit #2 Tests:

		Coal As Received(4)			SO_2			Particulate			NO_x	
Date	Gr MW	% Sul	% Ash	Btu/lb	LB/HR	PPM	LB/MMB	LB/HR	LB/MMB	% OPAC	LB/HR	LB/MMB
10/76	331	0.56	7.96	8368	1309	178	0.42	88.6	.029	11(2)	917	0.30
11/76	327	0.59	7.86	8484	696	83	0.23	91.4	.029	10(2)	993	0.32
12/76	324	0.64	7.87	8690	780	98	0.25	105.7	.034	16(2)	784	0.25
3/77	335	0.63	7.96	8739	664	84	0.20	91.9	.029	8(2)	1399	0.38
6/77	305(8)	0.72	8.48	8929	596.5	67	0.189	104.2	.051	8.4(2)	1161.4	0.403

Notes:
1. NO_x emissions guaranteed by boiler supplier equal to NSPS.
2. Average EDC monitor opacity.
3. Qualified observer, EPA Method 9.
4. Based on percent moisture of coal for period prior to test.
5. SO_2 test done at 230 MW (two scrubbers on line).
6. Two scrubber modules on line, all other tests done with three modules on line.
7. SO_2, NO_x tests 340 MW, particulate tests 260 MW.
8. SO_2, NO_x tests 305 MW, particulate tests 198.5 MW.

COSTS

Actual construction costs for the two units with some portion estimated through June 1977 are listed in Table 5. The emission control system alone costs $63.98 per net kW. When the scrubber sludge disposal system is added, the cost rises to $83.74 per net kW. The bottom ash and other disposal ponds are not included in these costs. Work is still proceeding on the installation and these costs will increase. These costs describe a combined particulate-sulfur dioxide removal system requiring minimal chemical addition, installed as a new unit and ordered in 1972. As a matter of interest, we compared it to the $30/kW for a combustion particulate and SO_2 removal system on a new unit, published by EPA in Volume 37, No. 55 of the Federal Register, March 21, 1972 as a result of the Kennecott Copper case. Escalating this number from mid 1971 to Spring 1975, the approximate cost center of gravity of the Colstrip Project, gives $40/kW which is still about one-half of our actual costs.

SUMMARY

In-depth pilot studies formed the basis for the design of the closed loop combined particulate and sulfur dioxide removal system. Careful attention to engineering details and coordination was required from both the vendor and the architectural engineer to build the system. The generation plant staff was trained and obtained a thorough understanding of requirements that this type of air pollution control system placed upon plant operation.

ACKNOWLEDGMENTS

Portions of this paper appear in AIChE Paper No. 17e, August 1977, The Second Pacific Chemical Engineering Congress. It was jointly written by Bechtel Corporation, Combustion Equipment Associates, Arthur D. Little, Inc. and The Montana Power Company.

TABLE 5

FGD SYSTEM COSTS
TWO GENERATING UNITS

		$	$/Net kW*
1.	Scrubber subcontract and out of scope additions by prime contractor (includes foundation, field wiring, insulation, heat tracing, painting, piping to ponds, piping between modules and field distributables in addition to the scrubber modules, ductwork, equipment supply and erection)	$33,677,261	$51.02
2.	Owners' cost including interest at $2,792,000	4,374,000	6.62
3.	Subtotal as of December 1976	38,051,261	57.65
4.	Estimated cost for work in progress (cost responsibility under dispute)	4,178,499	6.33
5.	Fly ash ponds (temporary storage), fly ash slurry transport system and evaporation pond	12,722,131	19.29
6.	Incremental costs (auxiliary boiler, FD fans, start-up and auxiliary transformer)	317,520	0.48
7.	TOTAL**	$55,269,411	$83.74

*This cost does not include an additional claim submitted by the subcontractor. It also does not include a value for reduced plant capacity, although for many purposes, it should.

**Net rating 330 MW/unit; gross rating 358.4 MW/unit.

EXPERIENCE WITH LIMESTONE SCRUBBING SHERBURNE COUNTY GENERATING PLANT NORTHERN STATES POWER COMPANY

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ABSTRACT

This paper discusses the physical description, operational problems, system performance, and future design considerations for the limestone wet scrubber systems at Northern States Power Company's (NSP) Sherburne County Generating Plant Units 1 and 2. The scrubber systems are designed for both sulfur dioxide and particulate removal. Units 1 and 2 went into operation in May, 1976 and April, 1977, respectively. Fuel for the two-unit plant is low sulfur (0.8 percent) western coal.

The Combustion Engineering, limestone, two-stage wet scrubber systems at the Sherburne County Generating Plant were constructed after a pilot unit was tested in 1973 and 1974 at an existing NSP plant. Testing on the pilot unit resulted in numerous design changes, but did eventually confirm that these tail-end limestone scrubber systems could remove at least 50 percent of the sulfur dioxide while firing coal containing 0.8 percent, and 99 percent of the particulate matter, down to an emission level equal to 0.04 grains per dry standard cubic foot.

Operation of the full-size units has uncovered many problems that did not show up during the pilot testing. Many of these problems continue to hamper operation and require an extensive commitment of manpower by NSP to sustain reliable unit operation.

EXPERIENCE WITH LIMESTONE SCRUBBING
SHERBURNE COUNTY GENERATING PLANT
NORTHERN STATES POWER COMPANY

INTRODUCTION

Northern States Power Company (NSP) serves energy customers in Minnesota, Wisconsin, North Dakota, and South Dakota. NSP employs approximately 6000 people and has a maximum generating capability of approximately 5500 megawatts (MW).

Construction of the two-unit Sherburne County Generating Plant (SherCo) began in August of 1972. Units 1 and 2 went into commercial operation in May, 1976 and April, 1977 respectively. The two units are identical and are of the 700 MW size.

The present plant site is approximately 1700 acres and is located in a rural area of south central Sherburne County, approximately one and one-half miles southwest of Becker, Minnesota and approximately 42 miles northwest of Minneapolis. The site is bounded on the southwest by the Mississippi River.

The air quality control systems for Units 1 and 2 at SherCo are limestone, two-stage wet scrubbers, utilizing a rod venturi as the first stage (particulate removal) and a marble bed as the second stage (sulfur dioxide removal).

GENERAL PLANT DESCRIPTION

STEAM GENERATORS

The two Combustion Engineering steam generators are controlled circulation, single-reheat, balanced draft units having a primary steam flow of 4,985,000 pounds per hour. Superheater outlet pressure is 2640 psig at a temperature of 1007 F, with an inlet feedwater temperature of 487 F. The reheat steam flow of 4,501,000 pounds per hour is raised to an outlet temperature of 1005 F from inlet conditions of 597 psig and 636 F.

TURBINE-GENERATORS

The two turbine-generators are General Electric tandem-compound machines. The turbines are four-casing machines with single-flow high-pressure, double-flow intermediate-pressure, and two double-flow low-pressure sections.

The turbines are rated at 660,000 kw with steam conditions of 2400 psig, 1000 F, and 1000 F reheat, while exhausting at 1.5 inches Hg absolute, with zero makeup, and extracting for seven stages of feedwater heating, boiler feed-pump drive turbines, air preheating, and flue gas reheating. The turbines are designed for continuous operation at five percent overpressure, and at this rating (2520 psig) the actual capability is about 740,000 kw.

FUEL SUPPLY

The two primary fuels are sub-bituminous coals from the Colstrip and Sarpy Creek areas of Montana. Coal is presently supplied to the plant by unit trains, with a two-unit delivery frequency of approximately ten 100-car, 10,000-ton trains each week.

Typical coal analysis is 25-percent moisture, 9-percent ash, 0.8-percent sulfur, and high heating value of 8500 BTU per pound. The ash analysis is significant in the operation of the air quality control systems (scrubbers). Typically, the calcium oxide reported in the ash is 17-percent.

The secondary fuel is No. 2 fuel oil, which is used in the main boiler ignitors and the auxiliary boilers.

AIR QUALITY CONTROL SYSTEMS (SCRUBBERS)

The two Combustion Engineering scrubber systems are tail-end limestone, negative pressure units. Each system consists of twelve scrubber modules, eleven of which are required for full load operation. The flow through each of the twelve scrubber modules is approximately 200,000 actual cubic feet per minute (acfm) of flue gas. The approximate inlet flue gas conditions are 310 F, 3.0 grains per dry standard cubic foot (gr/dscf) particulate, and 700 parts per million (ppm) sulfur dioxide (SO_2). The approximate outlet flue gas conditions are 180 F, 0.04 gr/dscf particulate, and 300 ppm SO_2.

SCRUBBER SYSTEM PROCESS DESCRIPTION

The function of the SherCo scrubber system is two-fold: to ensure that the particulate matter in the flue gas leaving the scrubber system does not exceed 1% of that entering or 0.04 grains per dry standard cubic foot (gr/dscf), whichever is greater; to ensure that the sulfur dioxide (SO_2) in the flue gas leaving the scrubber system does not exceed 50% of that entering or 200 parts per million (ppm), whichever is greater. This will bring the total emissions to a level which will satisfy the requirements of the Minnesota Pollution Control Agency. The limits for the SherCo stack emissions are: 0.96 lb/MBTU SO_2 and 0.087 lb/MBTU particulate.

FLUE GAS FLOW

The scrubber system process begins as the flue gas leaves the air preheaters at approximately 310 F. The flue gas is channeled to the scrubber inlet ductwork which is designed to permit a selection of gas flow paths while assuring that the velocity in each duct is high enough to maintain the particulate matter (fly ash) in suspension and attain equal distribution to the scrubber modules in service. This is achieved by providing individual (laned) ducts with turning vanes at the inlet to each scrubber module.

A single blade shutoff damper at the inlet of the duct to each scrubber module is used to isolate the module. After the flue gas passes the inlet damper, it enters the scrubber module (Figure 1) at the scrubber first stage (also known as the primary contactor). The scrubber first stage is a venturi section of ductwork with two parallel rows of horizontal rods in the throat of the venturi. These rods are perpendicular to the flue gas flow. A slurry spray is introduced just prior to these rods. Some of the slurry is sprayed into the gas stream and is carried to the rods by the gas flow. The remainder of the slurry is sprayed on the walls of the venturi section to prevent deposits from accumulating on the surface of the wall. The spray originates in the scrubber module reaction tank, and is pumped to the first stage spray nozzles by the spray water pump. The purpose of the rod section is to reduce the cross-sectional area of the duct and thus cause a rapid increase in the flue gas velocity. This increased velocity causes the entrained particulates (fly ash) to be driven into the liquid droplets of the slurry supplied by the first stage spray nozzles, and thus be captured by the droplets.

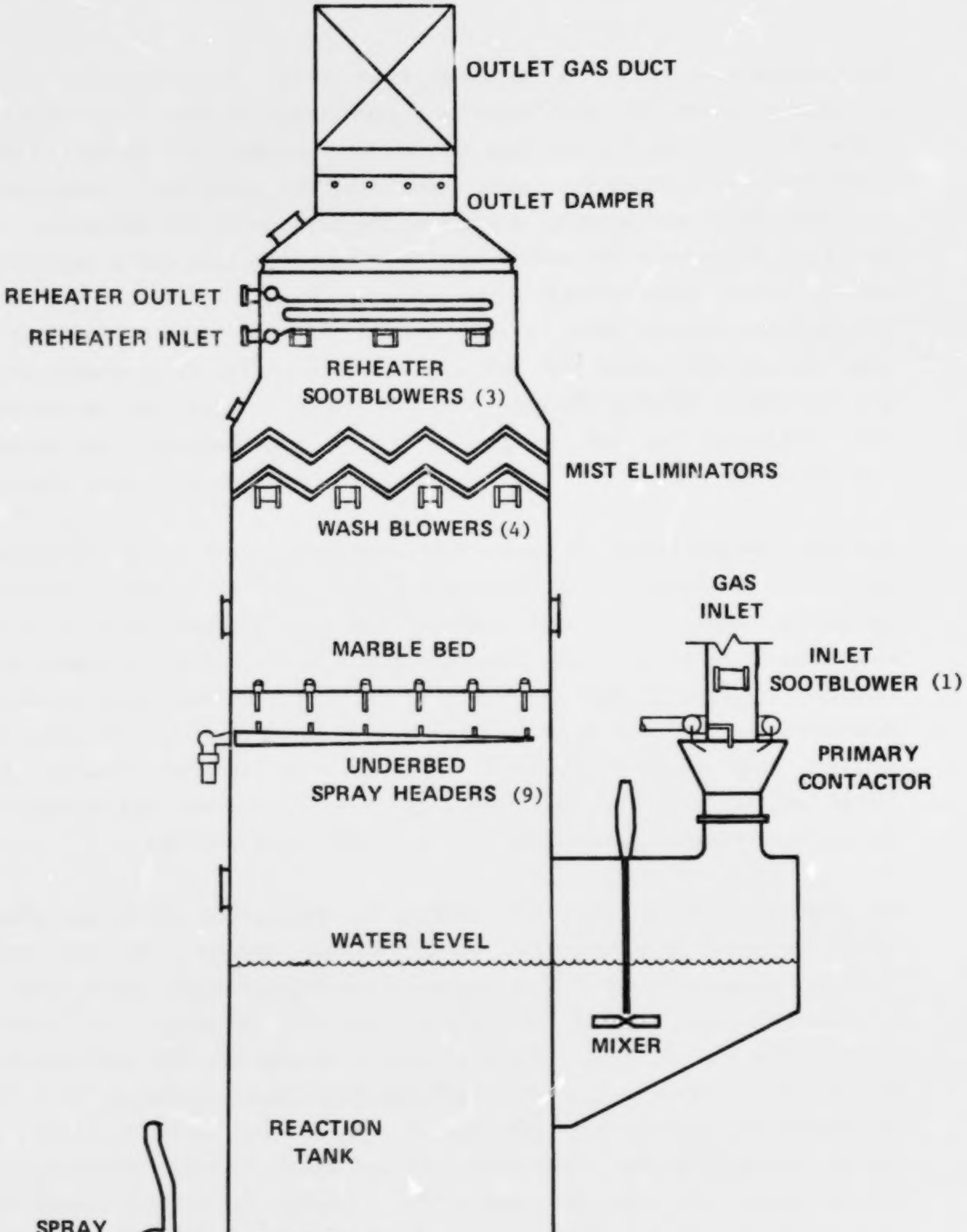

Figure 1

Scrubber Module.

The flue gas that leaves the scrubber first stage is relatively free of particulate matter, but still contains a significant portion of the original sulfur dioxide. The flue gas then travels into the main part of the scrubber module where it contacts the ladder vanes under the marble bed. These vanes help straighten and equally distribute the flue gas to the marble bed. As the flue gas contacts the marble bed, it is wetted by both the slurry on the bed and the slurry constantly being sprayed to the bed by the underbed spray nozzles. The underbed spray nozzles receive the slurry from the spray water pump. As the flue gas passes through the marble bed, it is thoroughly mixed with the slurry entering the bed. While traveling through the marble bed, some additional fly ash is captured by the slurry; however, the primary function of the marble bed is to complete the sulfur dioxide removal process.

When the flue gas leaves the marble bed, approximately 99% of the particulate matter and at least 50% of the sulfur dioxide have been removed. The flue gas is now cleaned, but still contains entrained moisture which is of the same composition as the slurry from the marble bed. In order to remove this moisture, the flue gas next contacts the mist eliminator (demister) section. This two-stage chevron shaped, fiberglass demister separates the entrained moisture from the gas to prevent plugging in the scrubber reheater. The liquid collects on the demister vanes, runs down the chevron, and collects at the low points where the droplets fall back into the marble bed.

The temperature of the flue gas leaving the demisters is 131 F, and almost all of the entrained moisture has been removed. However, the gas still contains enough moisture that condensation could occur which would have a detrimental effect on the induced draft (ID) fans and stack. To prevent condensation, the flue gas travels upward through a finned tube reheater section which increases the flue gas temperature by approximately 40 F. By increasing the temperature of the flue gas, the relative humidity of the flue gas is reduced, thereby preventing corrosion of the ID fans, breeching, and chimney liner. The flue gas leaves the reheater and travels through the outlet ductwork, to the ID fans, and then from the fans to the two outlet ducts and finally to the stack.

LIQUID PROCESSES

The raw limestone is ground in wet ball mills and delivered as a 4 percent slurry to each module's reaction tank (Figure 2). The slurry in the reaction tank contains the limestone, fly ash, calcium sulfate and dissolved ions (especially calcium, magnesium, and sulfate). The slurry is continually mixed in the reaction tanks by one vertical entry and two horizontal entry mixers. The slurry is then pumped, by a single spray water pump, to the first stage spray nozzles and the underbed spray nozzles, at flows of approximately 3,540 and 1,900 gallons per minute respectively. The slurry collects the fly ash and the sulfur dioxide from the flue gas and returns to the reaction tank.

The process of removing the SO_2 from the flue gas is achieved by using two additive sources; the calcium oxide in the fly ash and tail-end addition of limestone.

Calcium Oxide In The Fly Ash

The calcium oxide in the fly ash is very reactive. The chemistry of removing SO_2 consists of the following overall reactions:

(1) $CaO + H_2O = Ca(OH)_2$

(2) $Ca(OH)_2 + 2\ SO_2 + H_2O = Ca(HSO_3)_2 + H_2O$

(3) $CaSO_3 + SO_2 + H_2O = Ca(HSO_3)_2$

(4) $Ca(HSO_3)_2 + Ca(OH)_2 = 2CaSO_3 + 2H_2O$

(5) $CaSO_3 + 1/2O_2 = CaSO_4$

The calcium oxide (CaO) coming from the furnace with the fly ash is first hydrated, as in Equation (1). Removal of SO_2 in the marble bed depends upon the formation of calcium bisulfite [$Ca(HSO_3)_2$] by the reaction of calcium hydroxide [$Ca(OH)_2$], Equation (2), and suspended calcium sulfite ($CaSO_3$), Equation (3), with sulfur dioxide and water. In the reaction tank, soluble calcium bisufite is converted to insoluble calcium sulfite, Equation (4), and the sulfite is oxidized to sulfate ($CaSO_4$), Equation (5). These reactions account for the solid waste products as well as regeneration of calcium sulfite reactant for recirculating to the scrubber.

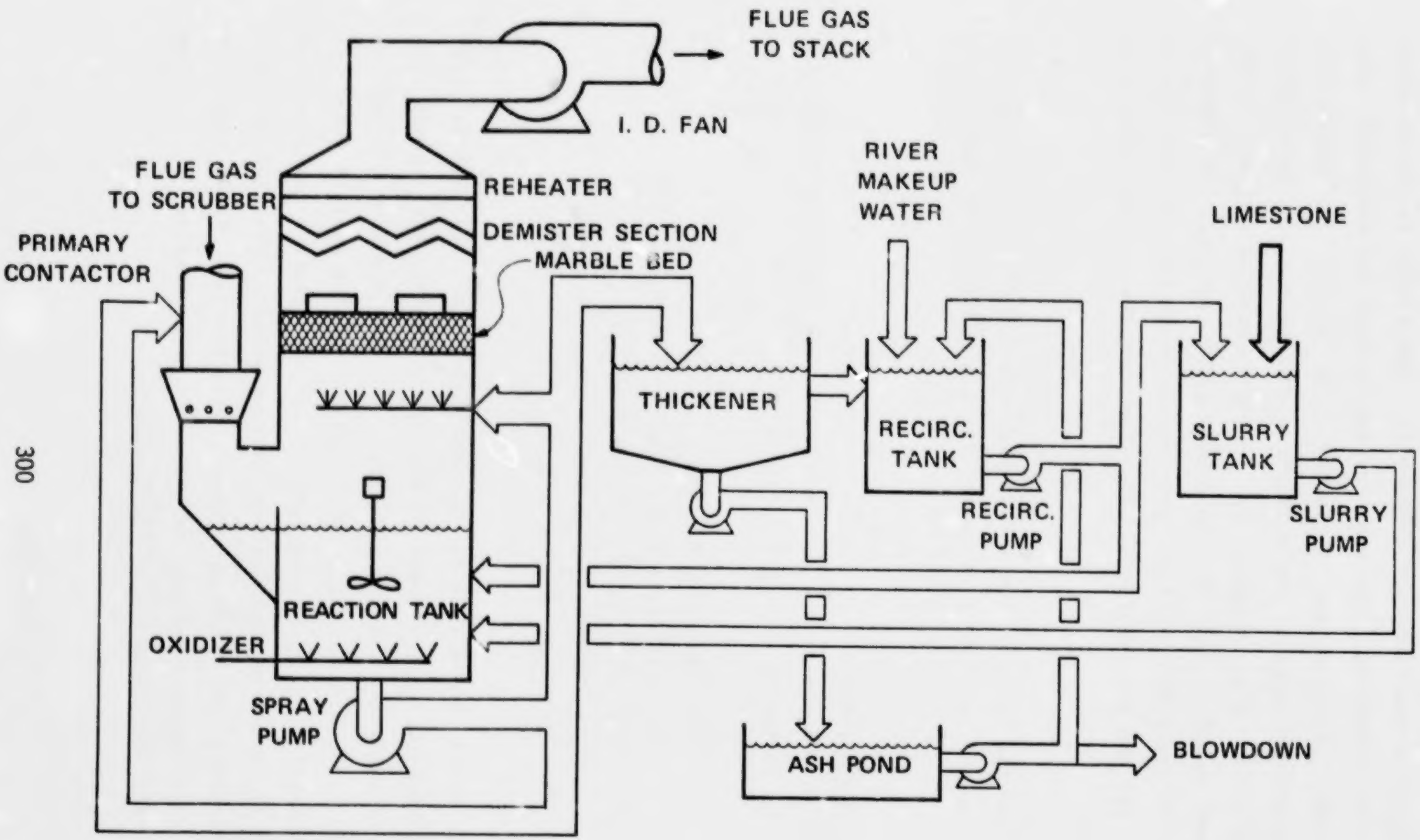

Figure 2

Sherburne County Wet Scrubber System.

Tail-End Limestone Feed

The process of removing SO_2 from the flue gas using limestone ($CaCO_3$) consists of the following overall reactions:

$$(6)\quad CaCO_3 + CO_2 + H_2O = Ca(HCO_3)_2$$

$$(7)\quad 2SO_2 + Ca(HCO_3)_2 = Ca(HSO_3)_2 + 2CO_2$$

$$(8)\quad CaSO_3 + SO_2 + H_2O = Ca(HSO_3)_2$$

$$(9)\quad Ca(HSO_3)_2 + 2CaCO_3 = 2CaSO_3 + Ca(HCO_3)_2$$

$$(10)\quad CaSO_3 + 1/2\ O_2 = CaSO_4$$

Equations (6), (7), and (9) are the principal absorption reactions. Sulfur dioxide reacts with relatively soluble calcium bicarbonate [$Ca(HCO_3)_2$] to form calcium bisulfite. In addition, solid calcium sulfite, recycled from the reaction tank, reacts with the SO_2 to form calcium bisulfite, Equation (8).

The oxidation reaction of calcium sulfite to sulfate, Equation (10), and the conversion of bisulfite to insoluble calcium sulfite, Equation (9), also account for the waste products as well as the regeneration of the calcium sulfite and calcium bicarbonate that are recirculated to the scrubber.

The calcium oxide addition from the fly ash and limestone additive system work simultaneously. Tests with and without limestone additive have shown that both additives contribute to the SO_2 removal process.

As the process of collecting fly ash and forming calcium sulfate continues, the percentage of solids in the slurry would continue to increase if left uncontrolled. In order to control the percent solids in the reaction tank, an effluent bleed-off flow (averaging 160 gpm) is drawn off the spray water pump discharge and sent to the thickener via the slurry transfer tank. This flow is automatically controlled by a nuclear density meter which senses the percent solids in the slurry and operates a control valve to maintain the solids at 10 percent. Ten percent solids has been determined to be the optimum concentration which minimizes scale formation. Scale formation is controlled by supplying sufficient (2 to 3%) calcium sulfate solids in the slurry which act as seed crystals and force the continually forming calcium sulfate to form on the seed crystals. At 10 percent total solids, the amount of calcium sulfate seed crystals is about 2 to 3% (the remainder of the solids are fly ash).

If the total percent solids were too low, the concentration of calcium sulfate seed crystals would drop and scaling would occur. If the total percent solids were over 10%, no major immediate adverse affect would be noticed; but with the excess solids in the slurry, the deterioration and wear of the slurry handling equipment would be accelerated.

The amount of calcium sulfate seed crystals formed depends upon the degree of oxidation from sulfite to sulfate. In order to ensure that all the sulfite formed is oxidized to sulfate, air is introduced into the reaction tank. This air enters near the two horizontal entry mixers, which ensures that the air is mixed with all the available sulfite. The sulfite is oxidized to sulfate, which acts as seed crystals.

The effluent bleed-off is discharged to the thickener where the solids are concentrated and pumped to the fly ash pond. The decanted liquid at the top of the thickener is returned to the scrubber recirculation tank where it is used as makeup water to the scrubber modules, flush water for slurry lines, and mist eliminator wash water.

The slurry in the fly ash pond is allowed to settle. The solids remain in the pond and the liquid is collected and returned to the scrubber recirculation tank.

SCRUBBER SYSTEM COMPONENT DESCRIPTION

Each scrubber system consists of twelve (12) scrubber modules and associated equipment. The 12 modules are arranged in a four by three matrix and located inside the Scrubber Building (between the air preheaters and the stack). This section will describe some of the major components of the scrubber system.

SCRUBBER MODULE

Externally, a scrubber module consists of three major components; the reaction tank; the shell; and the first stage. Internally, a scrubber module consists of many components, some of the major ones being: the scrubber first stage rods and nozzles, the underbed distribution ladder vanes, the marble bed, the mist eliminators, the reheater, the reaction tank mixers, the underbed spray nozzles, the mist eliminator wash blowers, and the inlet and reheater soot blowers.

SCRUBBER SHELL

The scrubber shell is the section of the scrubber module from the top of the reaction tank to the top of the flue gas reheater. The scrubber shell is constructed of 1/4 inch carbon steel plate which is internally coated with a Ceilcote flake fiberglass lining, up to the reheater elevation. The purpose of the coating is to protect the carbon steel from corrosion caused by the slurry.

REACTION TANK

The reaction tank is located directly below the scrubber shell section. The tank is rectangular in shape and has a capacity of approximately 66,000 gallons of slurry which provides about 12 minutes retention time for the slurry before it is pumped through spray water piping to the spray nozzles. The reaction tank is constructed of 5/8" carbon steel plate which is not coated below the slurry level.

SCRUBBER FIRST STAGE

The scrubber first stage section (Figure 3) attaches to the side of the reaction tank and connects it to the inlet flue gas duct. The first stage consists of: an inlet section, a skirt section, a venturi section, and a holding tank which is the connection to the scrubber reaction tank.

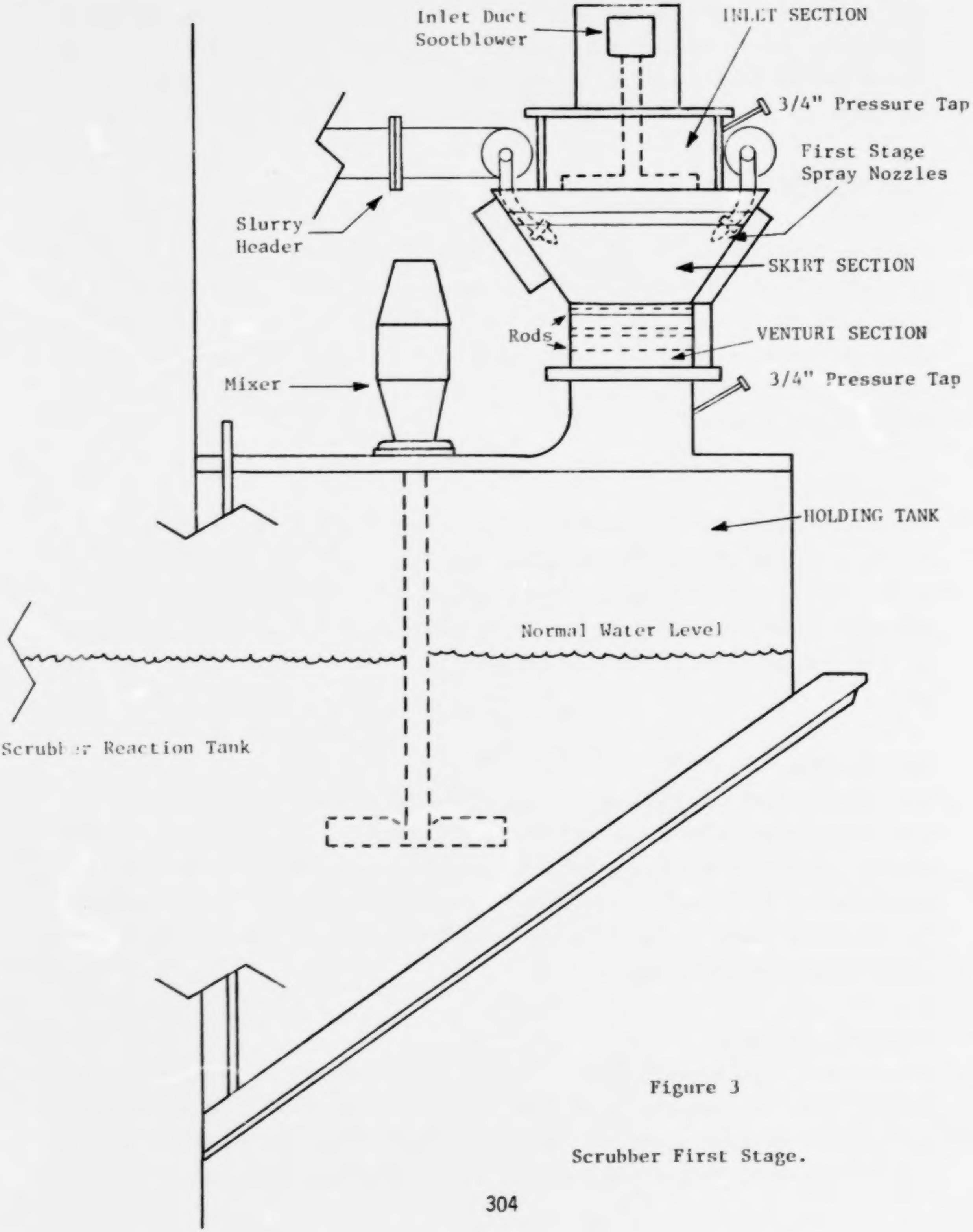

Figure 3

Scrubber First Stage.

Scrubber First Stage Inlet Section

The first stage inlet section is a 3'-0" by 25'-6" rectangular section of ductwork which connects the flue gas inlet duct to the remainder of the first stage. The inlet section is constructed of 1/4" carbon steel plate and included in this section is the scrubber inlet soot blower which is designed to remove any deposits created at the wet-dry interface.

Scrubber First Stage Skirt Section

The first stage skirt section houses the first stage nozzles and is the connection between the inlet section and the venturi section. The skirt section is constructed of 316L stainless steel. Twenty-eight first stage spray nozzles penetrate the top of the section around the perimeter of the top plate.

Scrubber First Stage Venturi Section

The venturi section is constructed of 316L stainless steel and houses the rods used to contact the flue gas and the slurry. There are two rows of rods which are constructed of 316L stainless steel. The spacing between the two rows of rods is adjustable by moving the bottom row up or down. This adjustment is accomplished by manually operating the scissor jacking system which is attached to the bottom row. The upper rods are welded to bars which are permanently fixed in the duct. The rods are assembled in sections approximately three feet by five feet, five sections in each venturi.

The rods are 2-1/2" O.D. schedule 40 stainless steel pipe. The rods connect across the width of the duct and are about 2'-2" long.

Scrubber First Stage Holding Tank

The holding tank is a combination of rectangular and triangular sections with the rectangular section attaching to the aforementioned venturi section. The holding tank is constructed of 1/4" carbon steel plate and houses the first stage mixer which is vertically mounted on the top of the tank.

SCRUBBER MARBLE BED

The marble bed consists of a perforated plate, marbles, and a slurry drainage system to the reaction tank. The marble bed perforated plate is 26'-6" by 18'-5" and is constructed of 316L stainless steel. The plate has an open area of approximately 40%, which is provided by 3/8 inch holes on 3/4 inch centers. The marbles are 13/16 inch diameter glass spheres situated on top of the perforated plate to a depth of approximately 4 inches. The slurry drainage system is provided by forty-eight drain pots located on the marble bed. The overflow pots are weir type drainage devices which are evenly spaced throughout the marble bed such that each pot drains approximately 10 square feet of bed area. The pots are constructed of a fiberglass reinforced polyester material. A domed, expanded metal stainless steel cover is attached to each pot to prevent the marbles from leaving the bed and entering the reaction tank.

MIST ELIMINATOR

The mist eliminator consists of two levels of vane assemblies located 10.5 feet above the marble bed. There is a vertical space of approximately four inches between the two levels of vanes. One hundred and forty-four vane assemblies are used in each scrubber module. Each vane is molded from a fiberglass reinforced polyester material.

FLUE GAS REHEATER

The finned tube reheater section is located approximately 10 feet above the mist eliminator section. It consists of 45 parallel carbon steel tube circuits arranged in horizontal rows in such a way that each row is staggered in relation to the row above and below. Each tube circuit (1-3/4" O.D. tubing) has three 180 degree bends, allowing four tube passes through the flue gas stream.

SPRAY WATER PUMP

One spray water pump is furnished for each module. The pump takes slurry from the reaction tank and discharges the slurry (5600 gpm) to the first stage spray nozzles (3540 gpm), to the underbed spray nozzles (1900 gpm), and the remaining 160 gpm is diverted to the scrubber sludge system (thickener) to maintain the proper percent solids in the reaction tank. Ni-Hard #1 (hard metal) is used in the pump casing, wear plates, and impeller.

OPERATING EXPERIENCE

SherCo Unit No. 1 was initially placed in service on March 16, 1976 and was released for commercial operation on May 1, 1976. SherCo Unit No. 2 was initially placed in service on January 25, 1977 and was released for commercial operation on April 1, 1977. For the operating period through August, 1977, the scrubber systems availability has been only 92.0% and 92.8% for Units 1 and 2 respectively (TABLE II). The scrubber system availability has increased slightly (1977 vs. 1976) as a result of completion of several design modifications and through additional operating experience with the system. Many problems still exist which are causing decreased availability, and increased manpower costs. A discussion of problems, past and present, is presented in the following sections.

STRAINER

The most significant problem affecting the overall scrubber system availability has been nozzle pluggage originally caused by failures in the Zurn duplex strainers on the discharge of the spray water pumps. The strainers were designed to remove solid particles greater than 1/4" in order to prevent plugging of the spray nozzles. Frequent pluggage of the strainers, mechanical failures, and bypassing of solids large enough to plug the spray nozzles resulted in a reduction of scrubber unit availability and increased maintenance requirements. The duplex operation of the strainers proved to be impractical to maintain and resulted in operation on a single strainer basket with a module shutdown required for cleaning. An extensive effort was put forth in an attempt to correct the problems; however, the efforts were not successful.

The duplex stainers were abandoned when Combustion Engineering installed new in-tank strainers. The new strainers consist of a large perforated semi-circular plate installed over the spray water pump suction with an oscillating and retracting wash lance for periodic backwashing. One strainer was placed in service in September, 1976 and seemed to result in a significant reduction in the number of plugged spray nozzles. Based on this successful operation, all modules on Units 1 and 2 were converted by March, 1977. Nozzle pluggage continues to be a problem, apparently caused by scale formation inside the perforated plate and piping headers which then breaks off and plugs the nozzles.

The experience, to date, with the system to clean the strainer perforated plate in unacceptable. Presently, as the perforated plate plugs, the blower is initiated which attempts to clean the plate. Complete cleaning is not accomplished which causes the plate t plug up again soon and require the blower again. As this continues, the slurry level in the reaction goes too high and the percent solids goes too low. The supply pressure and capacity to the blowers is the suspected problem and a new system for the supply is being designed.

MARBLE BED

Operation of the marble bed has been relatively satisfactory following a modification of the overflow pot covers. During the first few months of operation of Unit 1, several failures of the expanded metal stainless steel pot covers resulted in a loss of marbles to the reaction tank with subsequent pluggage of the spray water pump strainer. The covers were annealed and no new failures occurred for several months. Since then, sporadic failures have occurred and wear is obvious on the covers. Some scaling and mud buildup have occurred on the marble beds, mainly in areas where spray nozzles have been plugged. Statements about long-term marble bed cleanliness can only be made after the spray nozzle/strainer pluggage problems are remedied.

SCRUBBER FIRST STAGE (PRIMARY CONTACTOR)

Deposits of mud at the wet-dry interface were initially a problem; however, modifying the steam blowers and increasing the blowing frequency have decreased this buildup. Deposits of mud on the rods have also been a problem. The original nozzles for this area extended into the gas stream to ensure that the rods were wetted and no buildup occurred, but the nozzles themselves collected deposits of mud. When these nozzles were replaced, the rods were not being covered with slurry and developed mud buildups. New nozzles, which cover the rods with slurry and do not extend into the gas stream, have been installed and have decreased this buildup problem.

Erosion of the scrubber first stage compartment and the rods themselves continues to be a problem. The addition of 316L stainless steel wear plates in the first stage compartment is successful in decreasing the compartment erosion problem caused by direct slurry impingement from the nozzles. Due to erosion, the existing rods (316L stainless steel) have not given satisfactory life. Rubber coated stainless steel rods were evaluated as a replacement, but the rubber came off in a short time. It should be noted that the first stage was originally designed for a pressure drop of 8-1/2 inches of water; however, in order to obtain guaranteed dust removal performance, the modules have been operating with a first stage differential pressure of 11-12 inches. One "test" module has been converted to a spray tower design, and is currently operating with a first stage pressure drop of 15-17 inches. The increased pressure drop has accelerated the rod and compartment erosion and the module has unsatisfactory demister cleanliness.

As a short-term "fix" to the rod erosion problem, NSP is in the process of welding stainless steel angle iron wear plates to the bottom row of rods in order to "buy time" until a long-term solution is proven.

REACTION TANK

Erosion of the sidewalls (above the water level) in the scrubber first stage hopper area has been the major reaction tank problem. This erosion is caused by the slurry cascading down from the first stage rods and impinging on sloped sidewalls under the first stage. The fiberglass flake liner in this area failed; therefore, 316L stainless steel wear plates have been installed in these erosion areas.

Settling of the slurry on the floor of the reaction tank has been an item of concern. This settling is caused by less than optimum mixing and inherent problems with total mixing in a rectangular tank. These deposits (approximately 6-8% by volume) are not considered detrimental to the overall operation of the scrubber system.

MIST ELIMINATOR

Mist eliminator pluggage continues to be a minor problem which requires periodic manual washing. Carryover of slurry through the mist eliminators is small as evidenced by the minor buildups on the downstream reheaters.

REHEATER

Four carbon steel reheaters experienced failures in the area of the return bends in 1976. Combustion Engineering metallurgical reports indicate that no internal corrosion existed and that external erosion had been a result of "washing" after an initial failure. The initial failures were attributed to weld failures caused by excessive stress at a point where a straightening bar was attached to the reheater tubes. Modifications are completed which removed the straightening bar in an effort to resolve the stress problem. Since then, only one minor reheater leak has occurred.

PIPING

Failures of the LaFavorite rubber lining, downstream of orifices, etc., have occurred in the main spray water and effluent bleed-off piping. When these failures occur, the rubber breaks off in chunks and plugs the downstream nozzles or headers. Such failures occurred to a small extent on Unit 1 and to a major extent on Unit 2. The rubber lining is being removed to prevent further header and nozzle pluggage: however, operation with the unprotected carbon steel sections has resulted in failures from erosion. A test program is currently underway to evaluate the performance of reinforced fiberglass, stainless steel, and rubber lined spool pieces.

SPRAY WATER PUMP

Present operation of the Worthington spray water pump indicates that the Ni-Hard impeller will require replacement about every 6,000 hours, and the Ni-Hard suction side wear plate about every 4,000 hours. Pump internals of 28% chrome-iron and rubber-lined internals have been installed on selected modules for evaluation.

COMMINUTER

The comminuter system was originally installed to grind up any large chunks which might collect on the reaction tank perforated plate. Operating experience with the scrubber system has shown that large chunks are not collected on the perforated plate and that scale eventually plugs the plate, therefore the lower portion of the perforated plate and the comminuters have been removed.

PROTECTIVE MODULE COATING

The Ceilcote lining of the Unit 2 scrubbers has had numerous failures. These failures have ranged from small "pinholes" to larger areas where chunks have actually fallen off. Since the Unit 1 scrubbers are not showing these failures and since NSP fears that these failures will continue, an evaluation program is underway to determine the cause of the problem.

MANPOWER REQUIREMENTS

The manpower required to maintain and operate the scrubber systems has been larger than anticipated. The largest unexpected manpower requirement has been the labor cleaning crews which are utilized to clean selected modules (rods, nozzles, headers, demisters, strainers, etc.) on a nightly basis.

A breakdown of the present manpower requirements is as follows:

16	Operators (4 per shift)
1	Electrician
4	Instrument and Control Technicians
2	Chemical Technicians
2	Engineers
21	Laborers (cleaning crew)
8	Maintenance
54	TOTAL

SCRUBBER SLUDGE DISPOSAL

The scrubber sludge disposal system (not provided by the scrubber manufacturer) which consists primarily of the thickener, the sludge pumps, and the associated piping and valving has encountered many problems. The major problems include the following:

1. Sludge pump suction piping and valve pluggage.
2. Dead-leg pipe pluggage.
3. Hydraulic and drainage problems with the sludge transport lines causing failures and freezing.
4. Electrical malfunctions of valve operators.

Due to the extensive problems encountered, a complete design review was performed and the following modifications will be accomplished.

1. Modify/relocate thickener sludge pump discharge piping.
2. Install slower operators on selected valves in sludge piping to reduce water hammer.
3. Install larger pumps and redesign piping in scrubber pump house.
4. Level the sludge lines to fly ash pond.
5. Install surge restraints (accumulators) on sludge lines.
6. Crosstie slurry inlets to thickeners (allow either unit to discharge into either thickener).

FUTURE CONSIDERATIONS

Northern States Power Company and Combustion Engineering are jointly sponsoring a betterment program for optimizing the performance of the Sherburne County scrubber systems. The objectives of this program are to:

1. Increase unit availability.
2. Reduce manpower and operating costs.
3. Develop a first stage section in order to achieve acceptable service life and/or increased particulate removal.
4. Improve instrument and control applications.

The specific items presently considered part of the betterment program are:

1. Effluent bleed-off system redesign.
   The present effluent bleed-off control valve will be removed and a separate pump will be installed on each module. This will stop the problem of control valve erosion and should reduce the number of large particles in the modules since the suction of the pump will be from the tank (before the in-tank strainer) versus after the strainer as presently exists. This concept has been approved and materials are on order to implement this on all 24 modules.
2. Effect of liquid to gas ratio (L/G) on SO_2 and particulate removal.
   This area is being evaluated and no action has been taken at this time.
3. Venturi-rod section redesign.
   A venturi-rod section, utilizing 4-1/2" rods and a different internal configuration, will be installed on module 110 to evaluate materials and design.

This program is expected to be completed by January, 1979.

In addition, NSP is purchasing a dedicated computer monitoring and control system for the scrubber area. The present computer management system, only monitors sufficient information to selectively schedule modules in and out of service to meet load requirements. Operating experience with the unit has established the need for a network of data monitoring and an operator interface in the main control room. The present system monitors only 3 analog and 26 digital inputs with no Cathode Ray Tube (CRT) displays. The new system will monitor 134 analog and 48 digital points and display the information on a CRT mounted in the main control room. Eventually, all routine system operations will be performed by the computer and existing logic systems. The new system is expected to be completed by early 1978.

NSP is currently researching different materials in many areas for possible future application. Some of these areas are:

1. Primary contactor rods.
2. Spray water pump internals.
3. Spray water piping.
4. Spray water nozzles.

NSP has recently signed a letter-of-intent with Combustion Engineering for two-stage scrubbers on Units 3 and 4 at Sherburne County, scheduled for operation in 1981 and 1983, respectively. These scrubbers will utilize a similar rod-venturi inlet section followed by a spray tower (rather than a marble bed). Future research work, utilizing the scrubbers on Units 1 and 2, will be performed to prove/modify the proposed design for the Units 3 and 4 scrubbers. This research work will be done on module 101 and will include the following:

1. A venturi-rod section, utilizing 7" rods and a slightly different internal configuration, will be evaluated for materials and design.
2. Effect of a Bulk Entrainment Separator (BES) on particulate removal.

   A BES (pre-demister) will be installed to evaluate if either demister cleanliness of particulate removal can be improved.
3. Effect of a wash tray on particulate removal.

   A wash tray will be installed to evaluate if either demister cleanliness or particulate removal can be improved.

OPERATING AND PERFORMANCE DATA

The following TABLES (I, II, and III) present operation and performance data for SherCo Unit 1 from May, 1976 through March, 1977, and both Units 1 and 2 for April through August, 1977. TABLE I presents the various scrubber system operating parameters based on one unit full load operation (11 and 12 scrubber modules in service). TABLE II details the scrubber system availability and the hours of boiler operation as well as the total megawatt-hours generated on a monthly basis. TABLE III presents the capital and operating cost based on the somewhat limited information available.

TABLE I

SHERBURNE COUNTY GENERATING PLANT
SCRUBBER SYSTEMS OPERATING DATA*

Limestone Feed Rate - LB./MIN	65-90
Flue Gas Flow Rate - CFM at 130° F	2,200,000
Differential Pressures - IN. WG	
Marble Bed and Mist Eliminator	6.0-6.5
Scrubber First Stage	13.0
Reheater	0.5
Ducts and Dampers	2.0
Gas Weight Entering Scrubbers - LB./HR.	7,445,000
Gas Temperature Entering Scrubbers - °F	310
Gas Temperature Leaving Scrubbers - °F	180
Reheater Duty - BTU/HR x 10^6	132
Solids to Disposal - LB./HR.	84,000
Makeup Water - GPM	2,000
Water Evaporated - GPM	625
L/G (Marble Bed) - GPM/1000 CFM	10
L/G (First Stage) - GPM/1000 CFM	17
Oxidizing Air - % Stoichiometric	350
Spray Water - pH	5-5.5
- Dissolved Calcium (ppm)	500-700
- Dissolved Magnesium (ppm)	1500-2500
- Dissolved Sulfate (ppm)	8,000-15,000
- Dissolved Sulfite (ppm)	0
- Solid Calcium (%)	10-15
- Solid Magnesium (%)	.5-1.5
- Solid Sulfate (%)	15-20
- Solid Sulfite (%)	0

* - Data based on full load operation of one unit (11 of 12 scrubber modules in service).

TABLE II

Sherburne County Generating Plant
Scrubber System Availability

DATE	GENERATION HOURS		MEGAWATT HOURS GENERATED		SCRUBBER SYSTEM AVAILABILITY (%)	
	UNIT 1	UNIT 2	UNIT 1	UNIT 2	UNIT 1	UNIT 2
MAY, 1976	657	--	318,050	--------	86	--
JUNE, 1976	688	--	372,450	--------	84	--
JULY, 1976	512	--	269,700	--------	84	--
AUGUST, 1976	705	--	421,110	--------	94	--
SEPTEMBER, 1976	566	--	349,470	--------	95	--
OCTOBER, 1976	606	--	385,610	--------	93	--
NOVEMBER, 1976	720	--	454,010	--------	93	--
DECEMBER, 1976	722	--	480,920	--------	95	--
JANUARY, 1977	642	--	386,060	--------	90	--
FEBRUARY, 1977	609	--	411,900	--------	91	--
MARCH, 1977	743	--	510,650	--------	95	--
APRIL, 1977	717	696	460,100	442,160	95	92
MAY, 1977	312	665	195,790	421,030	92	91
JUNE, 1977	248	720	127,290	423,680	92	90
JULY, 1977	736	602	431,910	382,890	97	97
AUGUST, 1977	687	674	401,960	366,500	96	94
TOTALS	9,870	3,357	5,976,980	2,036,260	--	--
AVERAGES	616.9	671.4	373,561	407,252	92.0	92.8

TABLE III

SHERBURNE COUNTY GENERATING PLANT

SCRUBBER SYSTEMS COSTS

ITEM	ACTUAL COST ($)	UNITIZED COST	ANNUAL REVENUE REQUIREMENTS ($)
I. Capital Cost	60,000,000	$41.6/kw	7,020,000 ①
II. Operating Cost			
A. 1976 (Unit 1 Only)			
1. Labor	778,616	.275 mills/kwhr	
2. Materials	331,391	.117 mills/kwhr	
B. 1977 (Through July 31, 1977)			
1. Operating (Labor and Materials)	840,491	.224 mills/kwhr	1,509,000 ②
2. Maintenance (Labor and Materials)	710,422	.180 mills/kwhr	1,218,000 ②
C. Limestone ③	650,000	.079 mills/kwhr	650,000
D. Energy			
1. Electrical (Pumps, Fans, Etc.) ③	1,125,000	.137 mills/kwhr	1,125,000
2. Fuel (Reheat Cost) ③	886,000	.108 mills/kwhr	886,000
3. Replacement Energy Cost		$13/Mwhr ⑤	2,109,000 ④
		TOTAL	$14,517,000

① Based on 11.7% as the annual fixed charge rate for pollution control equipment

② Annual estimate based on seven (7) months of actual operation (January through July, 1977)

③ Estimates - based on projected annual costs and generation for 2-unit operation [$.60/10^6 BTU; 65% capacity factor]

④ Based on actual megawatt-hours lost due to scrubber problems (Unit 1-January through August, Unit 2-April through August, 1977) - extrapolated to full year of operation

⑤ Average annual NSP replacement energy cost

SUMMARY AND CONCLUSION

The wet scrubber system at NSP's Sherburne County Plant was designed with the objective of achieving a minimum availability of 90% with removal efficiencies of particulate and sulfur dioxide which would meet emission limits. Combustion Engineering, the scrubber system supplier, gave performance guarantees which would meet the emission limits. The average scrubber availability (TABLE II) for both units has been approximately 92% and compliance with emission limits has been demonstrated in several tests.

NSP's current objectives are to improve the scrubber system availability to 97% while maintaining or improving particulate and sulfur dioxide removal, and to reduce the system operating and maintenance costs. Future development activities will include: applying an improved control (scrubber management) system; continuing to follow sludge utilization developments; and improving (or replacing) unsatisfactory mechanical or control components.

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SAARBERG – HOLTER FGD PROCESS: SECOND GENERATION LIME-BASED FGD SYSTEM

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ABSTRACT

This paper describes the application of the SAARBERG-HOLTER process in the flue gas desulfurization plant at Saarbergwerke AG's Weiher II power station.

The SAARBERG-HOLTER process is a wet lime-based scrubbing process, but with the use of certain additives, a clear scrubbing liquor is achieved and the familiar problems attached to customary first generation lime scrubbing are avoided--namely incrustation, plugging, regulation of the pH values, and sludge deposits.

Thus, the SAARBERG-HOLTER process is not simply a variation of conventional lime-based FGD systems, but a truly second-generation desulfurization process based on lime. Besides using the alkaline, clear washing liquor, which is the defining characteristic of the process, energy-saving gypsum is obtained through a fully integrated process step. The gypsum needs neither settling ponds nor chemical fixation and makes stockpiling possible.

Two further installations are currently being constructed which, when operational in the chemical industry and electricity generation, will be the largest FGD plants operating on a commercial basis in Europe.

SAARBERG-HOLTER FGD PROCESS

- Second generation lime-based FGD-System -

Michael Esche *

1. Introduction

As early as 1968 Saarbergwerke AG, a major integrated energy enterprise with West Germany's second largest output of hard coal, began research into appropriate technological methods of preventing atmospheric pollution. The insights and experience gained led to the decision in 1974 construct and operate a second and third plant for the desulfurization and purification of flue gases:

- 125,000 Nm³/h (≙ 40 MW) plant for the desulfurization of flue gas at the 150 MW coal-fired power station 'Weiher II'

- 30,000 Nm³/h (≙ 10 Gcal) plant for the purification of stack gas at a waste products incinerator to simultaneously remove HCl, HF and SO_2.

SAARBERG-HOLTER UMWELTTECHNIK GMBH (SHU) was entrusted with the development and technical realization of these projects and asked to design and validate a large-scale

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system to purify stack and flue gases compatible with the requirements of the task itself and the needs of industry.

The plants received financial support from the Federal German Ministry of the Interior and were set the aim of meeting the demands of commercial operation :

- Construction of a structurally simple desulfurization system without stand by units and avoiding the kind of complicated chemical plant technology undesirable in power stations
- Proven high reliability and an operational availability exceeding that of a power station
- High degree of desulfurization with low energy consumption over a wide range of boiler loadfactor
- Solving the "sludge"-disposal problems of the by-product
- Acceptable capital investment, labor and operating costs compared to other desulfurization processes

2. Process Chemistry

Over 80 % of the FGD plants in the world are wet-scrubbing systems based on lime or limestone. The individual processes only vary in the design of the scrubber because for the rest they nearly all employ a lime or limestone slurry as sorbent.

The SAARBERG-HOLTER process (SHU-process) is not a variant of conventional lime/limestone-based FGD systems but a truly second-generation process based on calcium. Besides the added advantage of obtaining gypsum in an integrated process step, its unique feature is the use of an alkaline but clear liquor for scrubbing. 1)

Process that not use a $Ca(OH)_2$ suspension very often lead to incrustation, plugging and scaling in the scrubber. Further disadvantages are the problems of pH-regulation and the relatively high energy consumption by the pumps which results from the solid matter contained in the washing fluid suspension. Moreover, unless special preparation plants are attached, the first-generation processes produce the by-product $CaSO_3$ sludge, the water content of which, because of its thixotropic properties, cannot be reduced below 60 - 70 % and consequently can only be stored in large sludge settling ponds. This environmental problem can often only be solved by costly chemical fixation.

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1) National and international patents pending or granted.

The SAARBERG-HOLTER process (SHU-process) is basically also a wet scrubbing based on lime, but the addition of certain additives to the slaked lime avoids the problems associated with conventional lime-based processes.
The clear washing fluid is achieved by adding small quantities of HCl to form highly soluble calcium chloride. The high solubility of calcium chloride produces a clear alkaline washing fluid instead of a $Ca(OH)_2$ suspension. This is the unique feature of the SHU-process. The high level of dissociation of the calcium chloride creates the required amount of calcium ions necessary to effect the bonding of the SO_2.

Somewhat oversimplified, the absorption of SO_2 by the washing fluid of conventional lime scrubbing occurs in accordance with the following chemical reactions:

$$Ca(OH)_2 + SO_2 \longrightarrow CaSO_3 + H_2O \tag{1}$$

$$Ca(OH)_2 + 2SO_2\ (+\text{acid}) \longrightarrow Ca(HSO_3)_2 \tag{2}$$

$$Ca(HSO_3)_2 + O_2\ (\text{air}) \longrightarrow CaSO_4 + H_2SO_4 \tag{3}$$

$$H_2SO_4 + Ca(OH)_2 \longrightarrow CaSO_4 + 2H_2O \tag{4}$$

$CaSO_4$ is present as $CaSO_4 \cdot 2H_2O$
$CaSO_3$ is present as $CaSO_3 \cdot 1/2\ H_2O$

In view of the simplified representation it should be noted that, for example, the formation of gypsum is only possible by means of an additional acid cycle and its introduction before and during the oxidation process while lowering the pH value to maintain the water-soluble and oxidizable $Ca(HSO_3)_2$.
With the SHU-process this additional acidifying step to reduce the pH-value is unnecessary due to the clear liquor and the low degree of proprietary additive concentration. The acid required for the reactions (2) and (3) is constantly released by the following reactions and at the same time reneutralized :

$$2HCl + Ca(OH)_2 \longrightarrow CaCl_2 + 2H_2O \qquad (5)$$

$$CaCl_2 + 2SO_2 + 2H_2O \longrightarrow Ca(HSO_3)_2 + 2HCl \qquad (6)$$

The integral production of gypsum, which is a feature of the SHU-process, is the inevitable result of the subsequent reactions (3) and (4).
The chlorine ions present in the system act as intermediate carriers via the calcium chloride and thus remain in the cycle and are not used up.

The use of the water-soluble additives has additional benefical effects with regard to pH regulation.
The scrubbing process no longer needs to be restricted to a pH value of 7 and to the narrow limits of ± 0.5 between scrubber inlet and outlet.
The washing fluid, drawn off at the top of the precipitation thank and re-used, can be adjusted to a

pH value between 7 and 12 at the scrubber inlet according to the degree of desulfurization required and the SO_2 content of the flue gas.

This variation of operating conditions in a conventional lime-based FGD process would certainly lead to extensive $CaCO_3$ formation, severe blockage and subsequent shutdown of the scrubber unit. The reason this does not happen with the SHU-process is that when the $CaCl_2/Ca(OH)_2$ solution - not suspension - reacts with the SO_2 of the flue gas in accordance with the equations (5) and (6), HCl is formed. As a result the formation of undesired $CaCO_3$ and $CaSO_3$ is suppressed while the HCl is neutralized by $Ca(OH)_2$. The pH value of the washing fluid falls of its own accord as a result of the SO_2 scrubbing to within the range 5 ± 0.5 which permits the formation of the soluble $Ca(HSO_3)_2$ necessary for the formation of gypsum as in reaction (3).

Whereas the advantages of an FGD system with a clear liquor process are obvious, questions often arise concerning the addition of chlorine ions, as hard coal in the USA and Europe often contains a certain amount of chlorine compounds. Also there are fears that the addition of chlorine ions could lead to considerable damage from corrosion.

These aspects do have to be considered when using chlorine ions if a highly concentrated chloride suspension - not solution (!) - is used and/or if the entire scrubbing process between inlet and outlet of the scrubber is conducted in a pH range of under 6.9 as is usually the case in conventional first-generation lime-based FGD processes.

For this reason, in the SAARBERG-HOLTER process, a secret water-soluble catalyst ABSORBEN 75 is introduced into the washing fluid in addition to the chlorine ions. ABSORBEN 75 causes reactions similar to HCl.
The most important function of ABSORBEN 75 is the controlled drop of the pH value of ca. pH = 11 at the scrubber inlet down to pH=5 at the discharge point of the scrubber before the oxidation stage. As a result of this "buffer" effect of ABSORBEN 75, corrosion which might otherwise occur is completely avoided. The pH values involved show this very clearly.
The inexpensive additive ABSORBEN 75 furthermore increases the amount of calcium available and permits a considerably higher degree of desulfurization.

A further advantage of the "clear liquor" process with the special additives, HCl and ABSORBEN 75, is that such a high number of calcium ions are present in the washing fluid that the L/G ratio can be reduced to the unusually low level for lime processes of 1.5 to 2 l/Nm^3 (10 to 14 gal/mscf).

This results in energy costs particularly for the water pumps which are far lower than those of other lime processes. In addition,not $CaSO_3$-sludge but high-grade gypsum is obtained in a fully integrated process step which only requires simple oxidation.

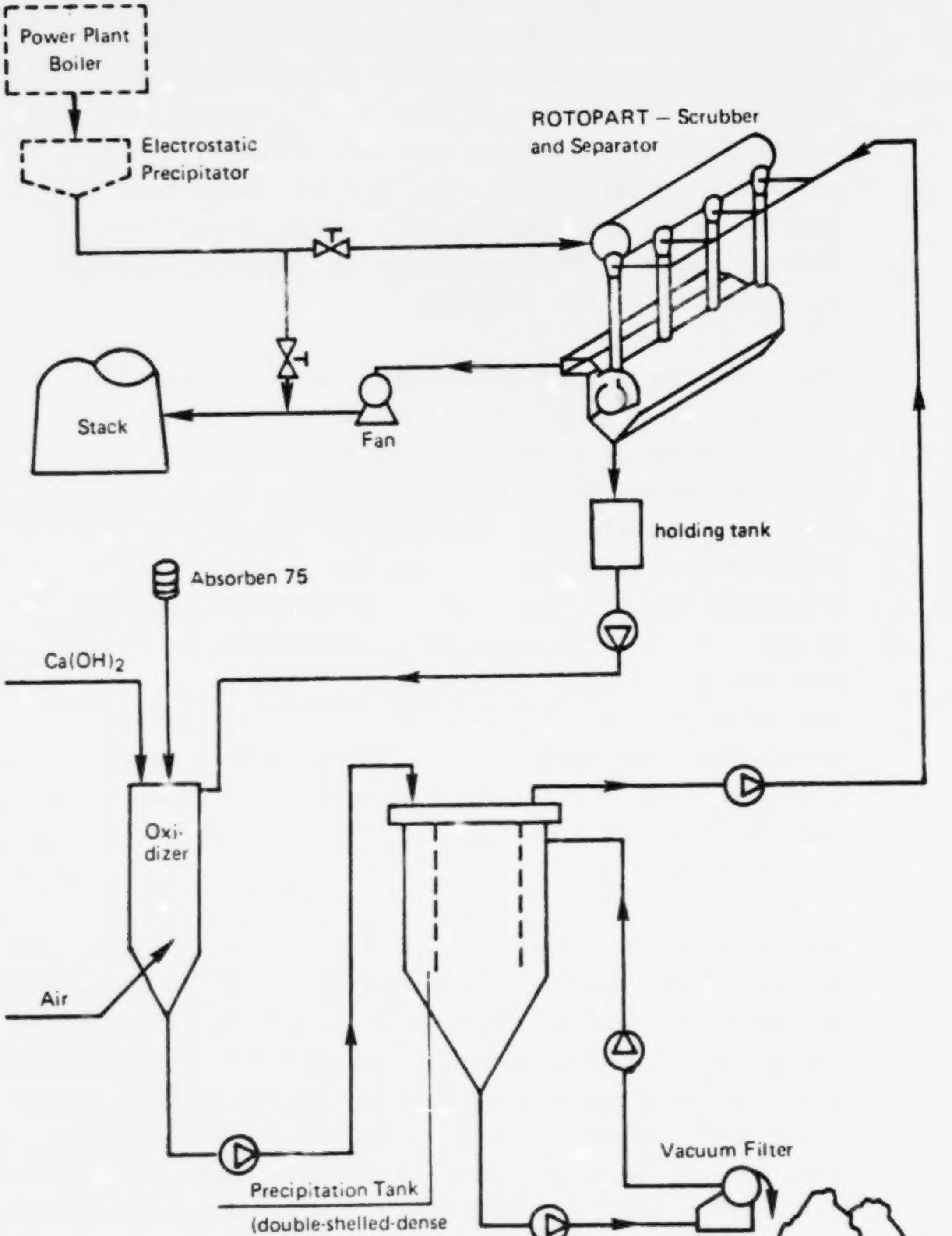
3. Process Description

The following FGD flowsheet shows the extremely simple design of the entire plant technology.

After the electrostatic precipitator the flue gases are fed into the vertically-aligned ROTOPART gas scrubber which is arranged in modular form. The modular form allows the plant to be adapted to any quantity of gas between 5,000 Nm³/h and 5,000,000 Nm³/h (3,000 - 3,200,000 scfm) without altering the special flow properties or the desulfurization criteria in the scrubber.

This modular design of the scrubber tubes was selected particularly as mass-transfer and gas distribution are far better than is often the case with large, less flexible scrubbing towers which have to be developed anew for the throughput capacity of each new power station.
In particular the relatively low mass-transfer values mean that in scrubbing towers large amounts of slurry have to be circulated in the scrubber. With the modular ROTOPART gas scrubber however it is not necessary to have slurry ciculating in the scrubber.

This has the advantage over conventional lime processes that there is no enrichment of residual particles and solid and soluble harmful substances, which in addition to the higher risk of errosion involves a technically extremely complex control system for the slurry especially in the scrubber.



Flowsheet for the SAARBERG–HOLTER FGD–installation for a flue gas volume of 125.000 Nm^3 (80.000 scfm)

In the ROTOPART-gas-scrubber* the flue gases are cooled in a single stage by means of jets arranged in one plane and simultaneously scrubbed by the special clear alkaline washing fluid. In addition to the injection nozzles the round scrubber tubes which are normally some 6 m (19,7 ft) long, have special water shedding rings** which prevent the washing fluid from running down the washing channels as an inactive film of water.

Following the scrubbing process, the flue gases from all the scrubber tubes are detached from the washing fluid by a single rotational particle ROTOPART-separator . This separator has no moving parts but works on the principle of the dense-medium separator in that the particles and SO_2-laden washing fluid are separated from the gas by centrifugal force created by diversion and flushed out **through** special water scraper rings.
For larger FGD plants too, this ROTOPART-separator consists of one single unit upon which the individual scrubber tubes are arranged in modular form one after another in a row like coke furnace chambers in a coking plant.

Not least of the advantages of the modular form was that it satisfied completely the demands of the power generating industry for high operational availability. Unlike many other FGD processes therefore the SAARBERG-HOLTER process requires no special reserve scrubbers or standby units. Additional demisting units to remove the most minute particles of moisture have turned out not to be necessary as the flue gases can be completely dried by the special ROTOPART demister.

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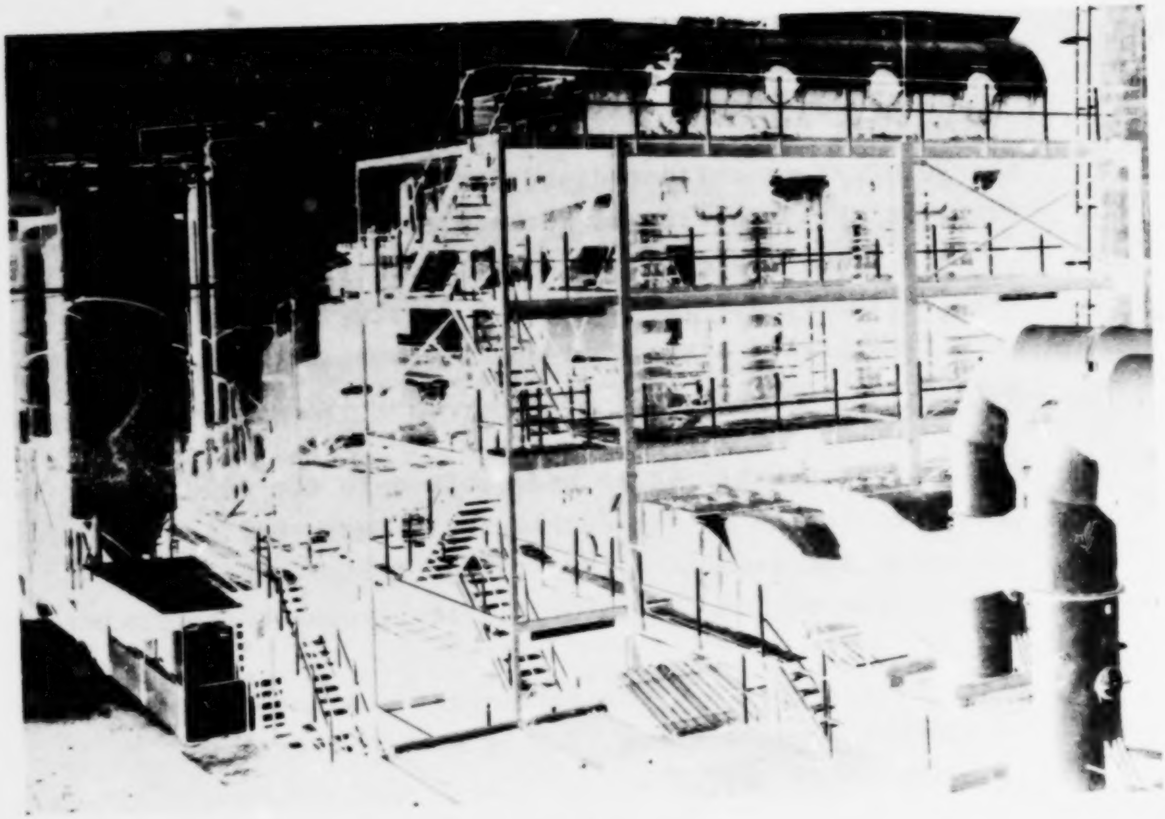
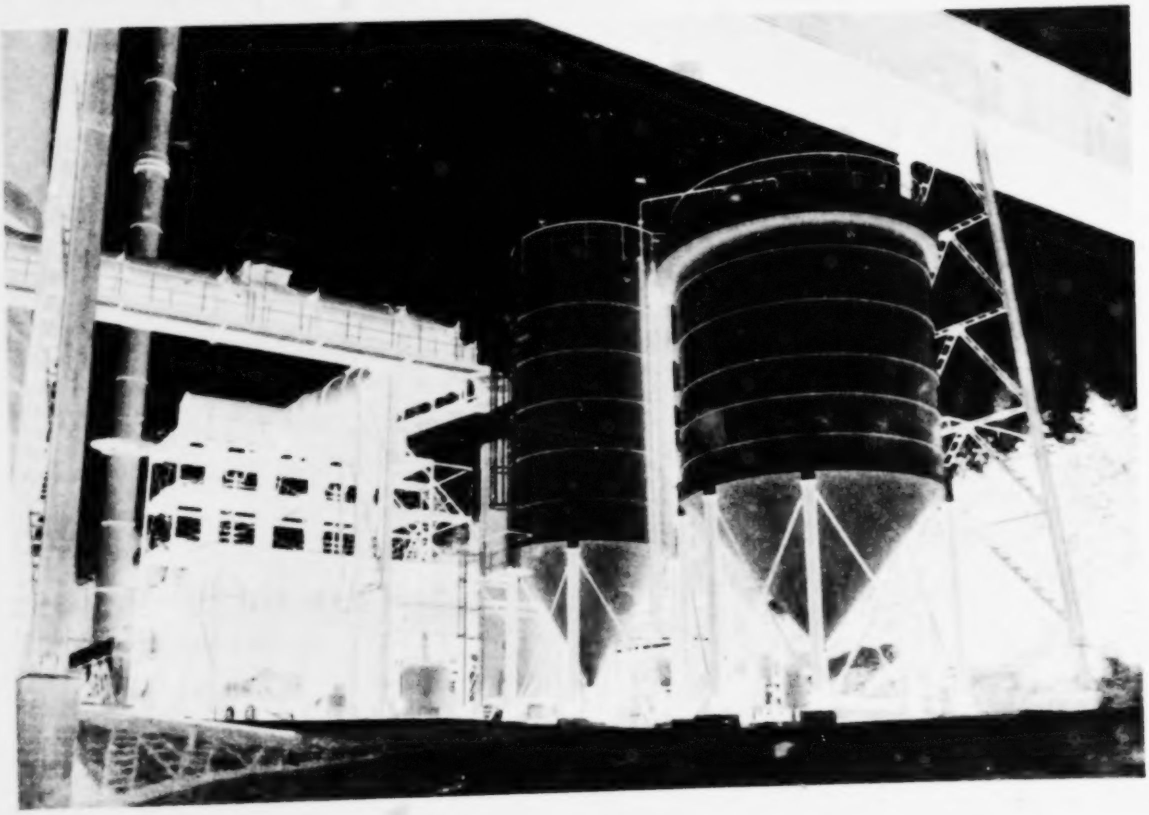
*
** International and national patents pending or granted



Saarberg-Holter FGD Plant „Weiher" Power Station

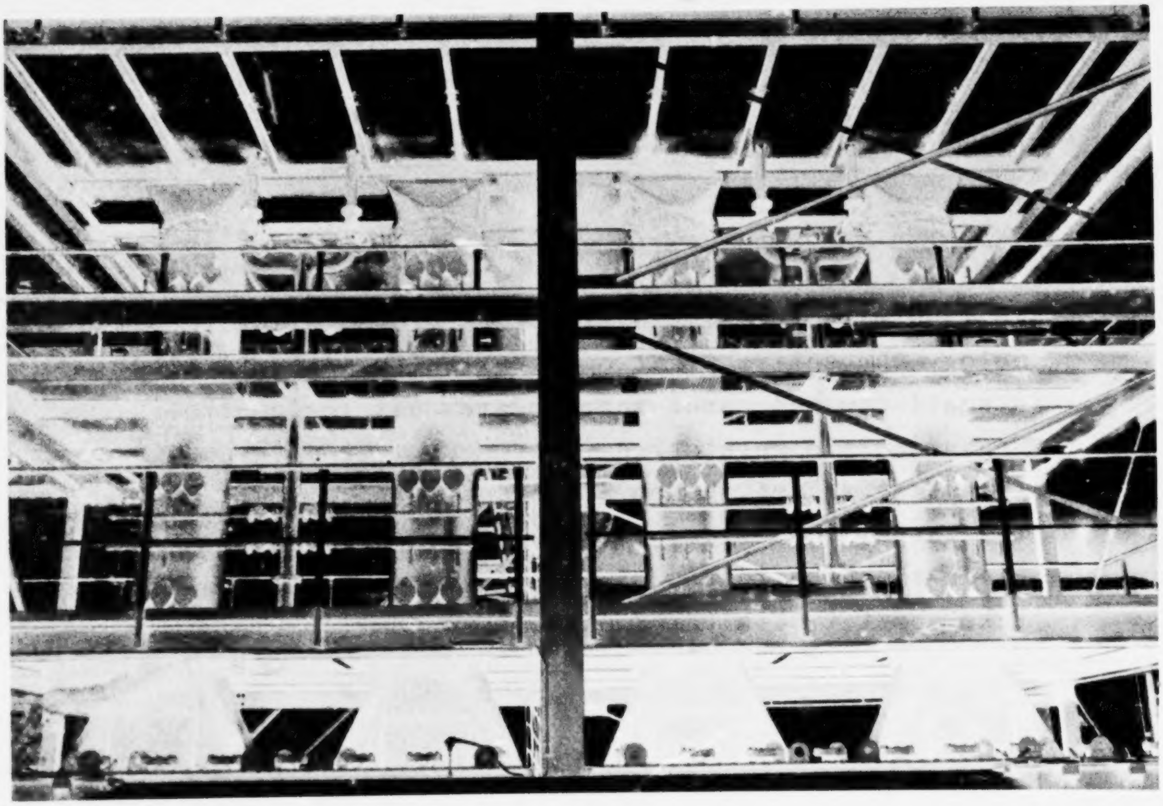


The washing fluid separated from the flue gas in the ROTOPART is collected and fed into an oxidation tank. In this tank air is introduced which converts it into a substance which consists of over 95 % $CaSO_4$ (gypsum) and less than 0.5 % $CaSO_3$. At the same time slaked lime is added to compensate for the loss of calcium ions in the gypsum formation.

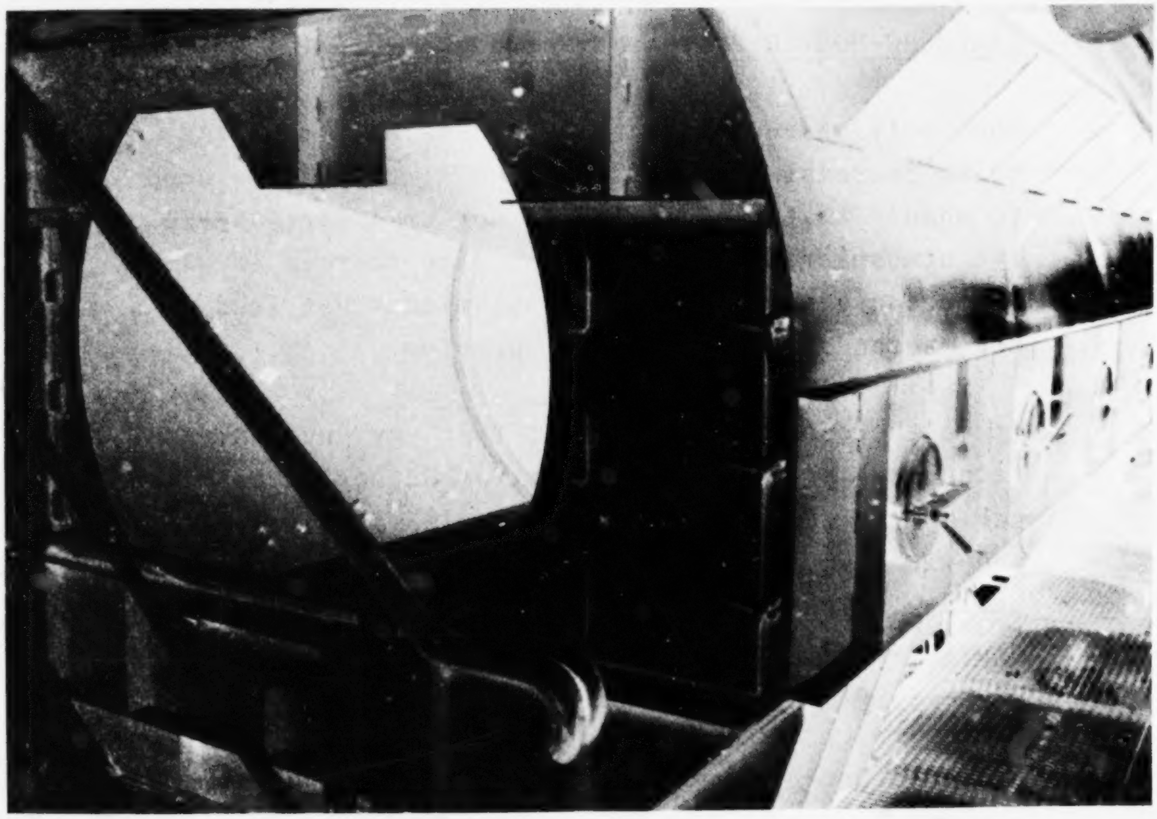
The gypsum crystals which have formed in the oxidation tank precipitate in the double-shell dense-medium separator and are fed into a vacuum filter. The filter reduces the residual moisture in the gypsum to approximately 20 %.
Tests have shown that the free water content can even be brought down to below 10 % by using thrust or swinging centrifuges.

At the top of the settling tank the clear washing fluid is extracted and re-injected into the scrubber, creating a closed washing fluid loop.
The SHU-process requires no equipment to remove waste water from the plant into the drainage system. Losses of water occur only with the saturation of the flue gases to saturation point and with the removal of the damp gypsum from the process.

The entire washing system is made of standard carbon steel, coated with a simple protective paint. The parts of the plant which come into contact with the desulfurized flue gas are provided with a protective plastic coating as sub-dewpoint temperatures are unavoidable.
It is expedient to provide a rubber lining for the I.D. fan where this is located behind the plant. Optimization studies have shown that the fan is better placed in front of the FGD plant. The installations currently being built are adopting this pattern.



Vertical Scrubber Tubes



Rotational Partial Separator (ROTOPART)

In line with the simple structure of the plant technology the entire plant operates fully automatically. A small control panel accommodates all the control, measuring and regulating instruments. From this room the plant is set in operation and stopped. Even with the larger installations this presents no problems given the high rate of load change. Both procedures take about two to three minutes.

4. SAARBERG-HOLTER Process By-product

When solving environmental problems in Europe and to an increasing extent in the USA, there is a need to ensure that the problem is not transferred from the atmosphere to the earth or into sources of water. The primary environmental problem must not lead to a new, secondary environmental problem.

It is therefore a requirement that any environmental technology employed to prevent atmospheric pollution has a by-product which can either be disposed of easily and without creating environmental problems or even better which can be re-cycled into the technological process. For lime-based FGD systems therefore, the production of a usable gypsum by-product in one integrated process step is an essential condition to qualify as a truly second-generation FGD process based on lime.

The SHU-process meets both these requirements by having as its by-product a gypsum which in a whole range of tests carried out by the gypsum industry has demonstrated its usability.

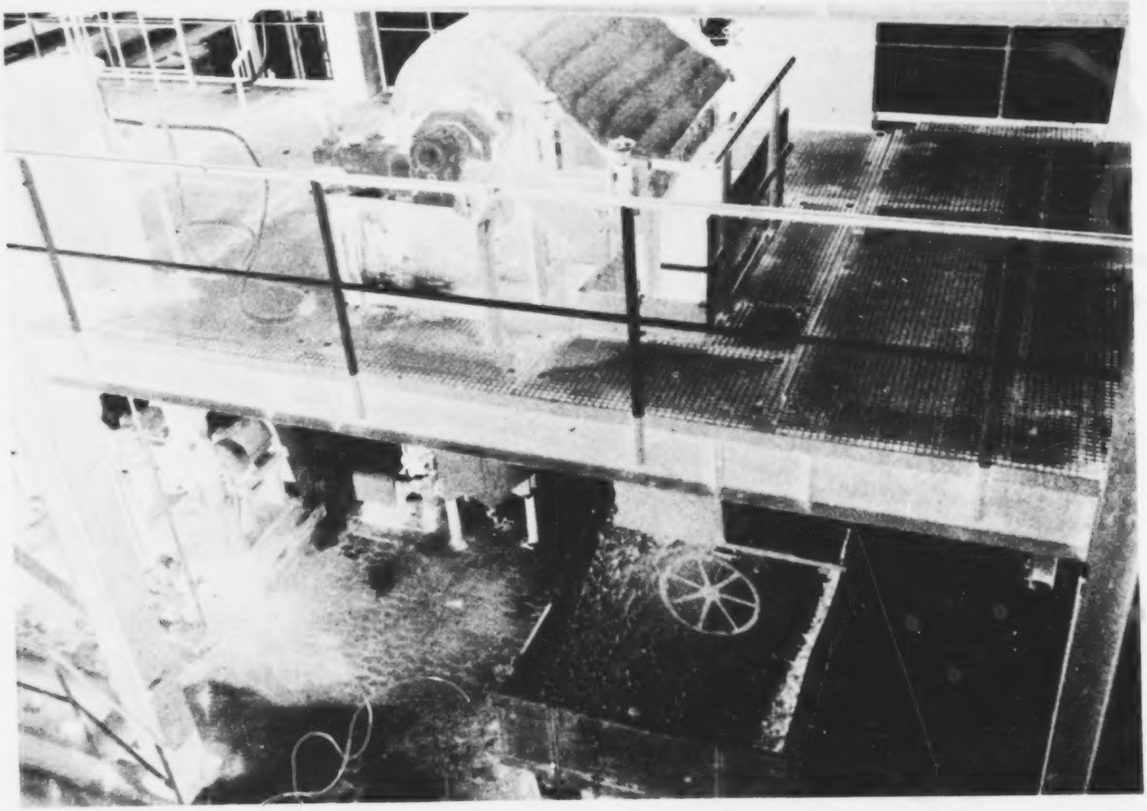
Unlike calcium sulfite sludge, gypsum can easily be de-watered so that the volume to be dumped is more than 50 % less than with calcium sulfite sludge. When the material is to be dumped the low water content facilitates transportation and allows the material to be piled thus considerably reducing the overall storage surface area required.

Furthermore dumped gypsum has no ecological impact since unlike $CaSO_3$ it is chemically entirely stable. Neither impermeable sludge ponds nor costly fixation is required.

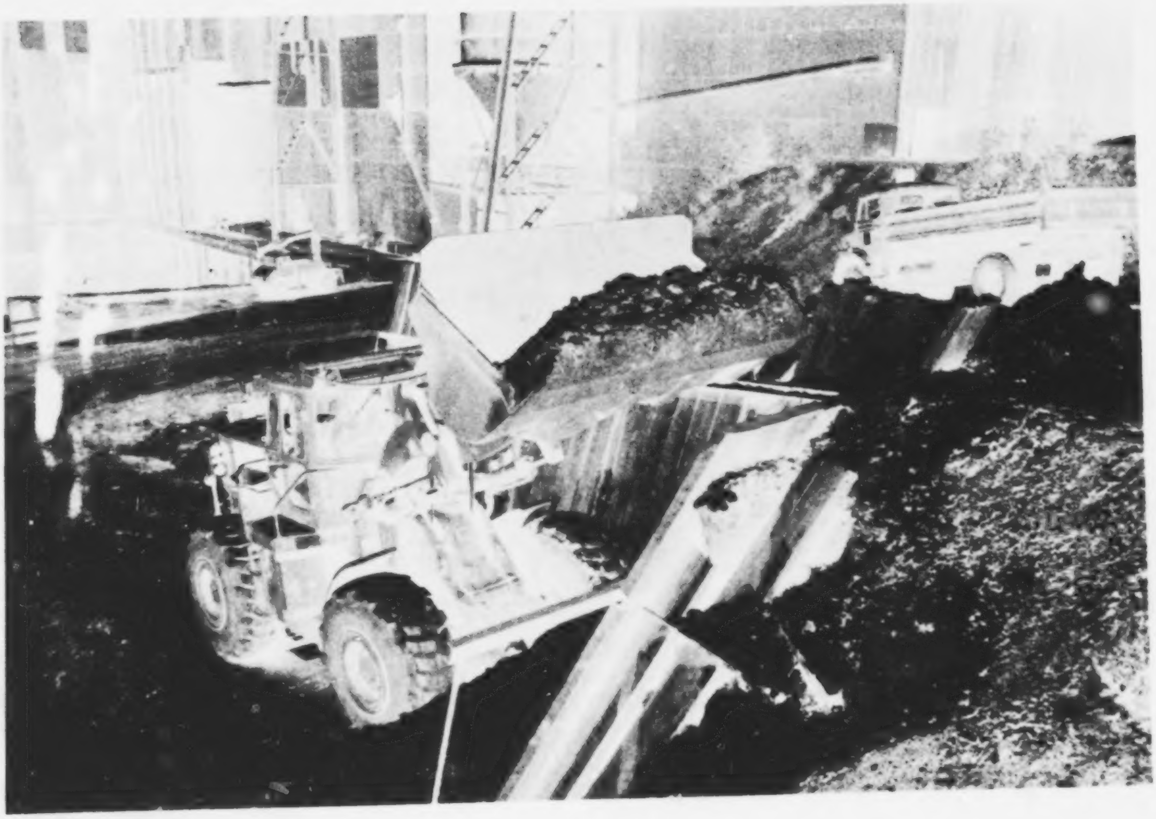
The gypsum produced by the SAARBERG-HOLTER process, in its dry form, is constituted as follows:

$CaSO_4 \cdot 2H_2O$	94 to 97 %
$CaSO_3 \cdot 1/2\ H_2O$	0.2 to 0.4 %
$CaCO_3$	0.1 to 0.5 %
Ballast material	5.7 to 2.1 %
	100 %

The ballast material consists of flue dust and the natural impurities in the lime.



Vacuum Filter



Stockpiled Gypsum

It is of course best if the gypsum by-product can be used rather than discarded. The prefabricated gypsum elements - for instance wallboards - produced from the power station gypsum in the SHU FGD plant are characterized by a specific gravity in the processed state which is some 10 % lower than that of boards made of mined gypsum but with approximately 50 % greater strength. According to the gypsum industry these results, obtained by the construction industry, are largely due to the higher degree of purity of this gypsum compared to natural gypsum and its particular crystalline form.
Investigations currently being conducted by TVA/EPA will show in which parts of the USA a byproduct gypsum could be used.

5. Operational Results with the SAARBERG-HOLTER Process

The following figures refer to the SAARBERG-HOLTER FGD system installed at the 150 MW coal-fired, medium-load power station "Weiher II" of Saarbergwerke AG.
The plant is d[illegible]ed to desulfurize a flue gas volume of 125,000 Nm^3 [illegible] 80,300 scfm).

a) Degree of Desulfurization

Depending on the coal used, the SO_2 content of the flue gases at a temperature of 140 °C ($\hat{=}$284° F) amounts to between 1,800 and 2,500 mg SO_2/Nm^3 ($\hat{=}$ 675 - 940 ppm). On the basis of an average of 2,200 mg SO_2/Nm^3 ($\hat{=}$ 825 ppm) in the flue gas and a liquid to gas ratio of 1.5 l/Nm^3 ($\hat{=}$ 10.4 gal/mscf) desulfurization levels of 90 % to 95 % are obtained.

As expected, with higher SO_2 concentrations in the flue gas, a higher liquid to gas ratio or a higher concentration of ions of the chlorine and particularly of the ABSORBEN 75 additive is required.

b) Scrubber reactant

As sources of calcium to absorb SO_2, either lime, limestone or calcium carbide can be used. In the flue gas desulfurization plant at the "Weiher" power station quick lime with over 95 % CaO is used, or alternatively calcium carbide which is available as a waste product in other industries.

Systematic test series have shown that in principle both products are equally suitable for the process. Calcium carbide has the advantage of lower cost as a waste product, but has disadvantages if the gypsum to be obtained as the by-product of desulfurization has to exhibit a particularly high degree of purity for special applications.

The consumption of calcium is in direct proportion to the degree of SO_2 concentration while the minute amounts of $CaSO_3$ and $CaCO_3$ in the gypsum by-product result in an average stoechiometric factor of 1.02 to 1.04.

Other scrubber reactants are hydrochloric acid and the additive "ABSORBEN 75 ". The quantities of these reagents consumed are - due to the cyclical nature of the process with their constant recovery by intermediate carriers - so small that they hardly have small effect on costs (6 - 7 $/hour for 750 MW with washing the filter). Coagulants are generally not required in the process.

c) Energy consumption

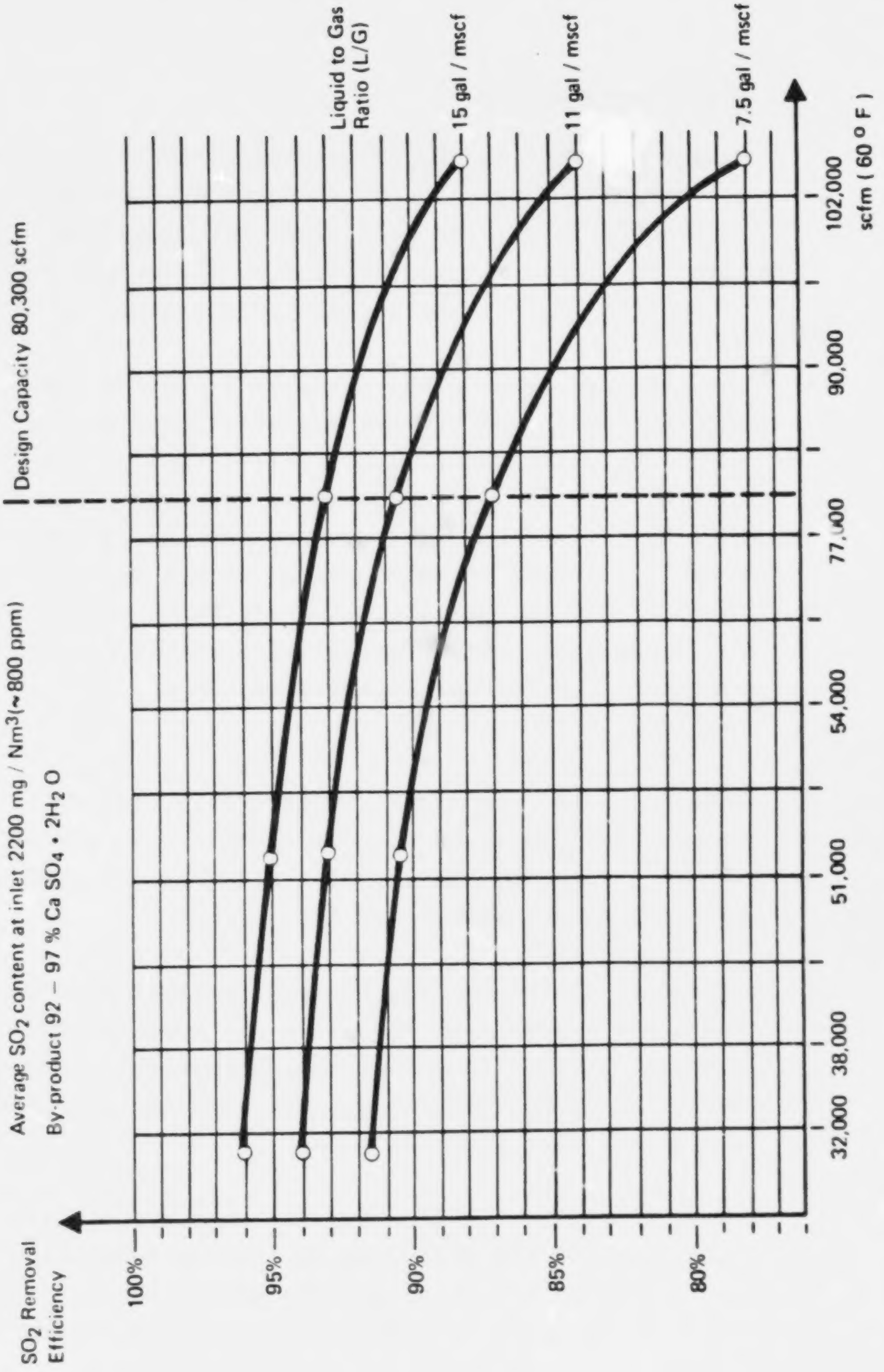
Energy consumption in wet-scrubbing FGD-systems is generally determined by the pressure losses from the scrubber and demisting unit, together with the liquid to gas ratio, i.e. the volume of washing fluid to be circulated per hour.

The pressure drop in the scrubber including the ROTOPART drying of the gas was originally in the magnitude of 300 and 350 mm (≙ 11.8 to 13.8 in) w.c. As a result of improvements and reductions of flow resistance, the pressure loss was significantly reduced to below 250 mm (≙ 9.8 in) w.c. Further improvements are currently being undertaken; for example laboratory experiments are in progress aimed at doubling the capacity of the scrubber sections.

Because of the low liquid to gas ratio, the total amount of fluid to be circulated of 180 to 250 m^3/h (≙ 105 - 145 cfm) and therefore, energy consumption by the pumps is relatively low.

Moreover, as the washing fluid is a clear solution and not a suspension, the pumps can operate without problems of erosion and at a high level of efficiency, so that compared with sludge pumps energy consumption is considerably reduced.

Total energy consumption for the entire FGD installation including all the pumps for the closed water loop and including the oxidation for the production of gypsum and including the vacuum filter amounts to approximately 0.95 % of the desulfurized power capacity.



Sulfur Removal Efficiency as a function of Flue Gas Volume

d) Flexibility

The FGD system can be operated within the range of 40 to 120 % of the layout capacity. If the L/G ratio remains constant, the cleaning of smaller flue gas volumes is initially improved, i.e. the level of desulfurization exceeds 95 % (see diagram).

e) Reliability and Availability

The SHU FGD-system at the Weiher power station was started for the first time in October 1974. The initial tests ran until 30th November 1974 when operating was suspended in order to allow the results and experience gained during the preliminary trials to be applied so as to maximize performence under the specific conditions prevailing in this plant. At the beginning of 1975 normal experimental running was resumed. Since then the plant has been operating successfully for over 17,000 hours.

During the first year, the plant was operational for approximately 90 % of the time. Disregarding the modifications and improvements as a result of experiments, the operational time amounted to approximately 94,9 %. Since April 1976, SAARBERG-HOLTER system availability has been 97.8 % in terms of boiler operational time (no stand-by units!).

Break-downs caused by the process itself due either to erosion, corrosion or to increasing scaling and plugging no longer occurred after, in addition to the utilization of the specified additives, the pH-regulator had been modified in April 1975, so that false pH-readings, which had led to erroneously stated pH-values, were eliminated. Mechanical defects in one of the washing fluid pumps were entirely satisfactorily eliminated by replacing face seals with packings.

f) Manpower for Operation and Maintenance

The manpower required for the plant is currently two men per shift, but the utility management assumes that with normal operation at the Weiher power station - without the special experiments and series of trials - economies of one man per shift could be achieved.

4 - 5 men per shift will be necessary to operate and supervise a desulfurizing plant for a 750 MW power station with a flue gas volume of 2,600,000 Nm^3/h (≙ 1,670,000 scfm), including current maintenance and supervising work.

The staff employed in the flue gas desulfurization plant does not need any particular skills due to the simplicity and easily comprehended nature of the process, which employs the technological devices mostly common to power stations and no innovatory of complex chemical techniques. Staff can therefore be recruited from within the power station.

g) Current Demonstration Model

After preliminary tests had shown that in addition to the desulfurization of flue gases from coal with a low sulfur content the gases from high sulfur coal (up to 4.5 %) could also be effectively dealt with using the SHU process, extended trials are now in progress to prove that these experimental results can also be attained over longer periods of time under normal operating conditions.
A commercial plant for the desulfurization of flue gas containing up to 5,000 - 6,500 ppm SO_2 is already under construction (see Chapter 6 b).

As mentioned above, such conditions require a proportionally higher L/G ratio. At the same time capital investment costs for a turnkey plant rise to an average of 70 - 75 $/kW, depending on the location of the generating station.

6. Plants Using the SAARBERG-HOLTER Process

Recently Saarbergwerke AG have begun to operate the first of the new generation of 700 MW coal-fired power stations in West Germany. The decision to build a further power station on this scale has already been taken. The FGD plant initially has to desulfurize a flue gas volume of approximately 650,000 Nm^3/h and will be expanded to 1,350,000 Nm^3/h respectively.

For both these projects the SAARBERG-HOLTER process represents an integral part of the coal-fired power generating station. The first to go into operation in 1978/1979 will be the first fully commercial FGD-plant of this size operating in Europe.

In addition enquiries have been received to construct FGD's for 5 other 700 MW major bituminous and subbituminous coal-fired power stations. Plans are further in progress for a FGD plant for a steam boiler fired by heavy oil with an 8%-sulfur content.

Following the successful operation of the SAARBERG-HOLTER FGD plant in power stations. the applications of this system to other sectors of industry have been considered:

(a) Following the construction of a small pilot plant of 3,000 Nm^3/h (≙ 1,900 scfm) in a waste products incinerator and the subsequent construction of the 125,000 Nm^3/h (≙ 80,300 scfm) desulfurization plant in the Saarbergwerke AG's Weiher II power station, in 1974, the construction of a demonstration plant for the purification of stack gas by simultaneously removing HCl, HF and SO_2 started in order to be able to use such plants in both waste product incinerators and specialized waste incinerators.

The plant is designed to treat the entire stack gas volume of the 5 t/h waste products incinerator and thus, purifies 30,000 Nm³/h (≙ 19,000 scfm) stack gas. As opposed to when the SAARBERG-HOLTER process is used in power stations, there is, besides the simultaneous removal of HCl, HF and SO_2, the special feature of higher stack gas temperatures on entry to the gas scrubbing system of 250 ° - 270 ° C (≙ 482 ° - 518 ° F).

In this plant the SHU process has shown itself to be entirely suitable to this area as well. While the removal of HCl and HF achieved levels of 97 - 98 % the removal of SO_2 reached levels above 90 % after modifications to the control mechanism which unlike in any power station had to deal with tremendous leaps in the level of concentration of 4,000 - 5,000 mg SO_2/Nm^3 (≙ 1,500 to 1,875 ppm) within the space of 2 to 5 minutes resulting in extremes of up to 10,000 mg SO_2/Nm^3 (≙ 3,760 ppm).

(b) In the area of the desulfurization of chemical plants, the SAARBERG-HOLTER system successfully developed in the power industry is now to be applied in a FGD plant which will handle 35,000 Nm³/h (≙ 22,000 scfm) of combined flue gases from two Claus ovens and a H_2SO_4-plant.

The design capacity for the max. SO_2-content is 13,500 mg/Nm³ (≙ 5.060 ppm). In the case of no flue gas coming from the H_2SO_4-plant the SO_2-content can rise to 17,500 mg/Nm³ (≙ 6,560 ppm). The sulfur removal efficiency had to be guaranteed to exceed 90 %.
The plant will go into operation in fall 1978.

(c) A further application of the SAARBERG-HOLTER process is the desulfurization of sintering belts in steelworks. An engineering study was carried out for 320,000 Nm³/h (≙ 205,600 scfm) plant with a SO_2 content of up to 2,500 mg/Nm³ (940 ppm) and particulate concentrations of up to 250 mg/Nm³ (0 0.1 grains/scf). A pilot plant of 3,000 Nm³/h (≙ 1,900 scfm) subsequently confirmed in practice the basic theoretical considerations and modifications to the process appropriate to this specific use with a degree of desulfurization of 90 - 94 %.

7. Summary

The SAARBERG-HOLTER (SHU) process has proved itself to be fully adapted to the demands of FGD plants for coal-fired power stations operating on a commercial basis. The system is conspicuously a truly second-generation lime process as it fulfills the necessary criteria and requirements without exception:

(a) Guaranteed high levels of desulfurization over a wide range of boiler loadfactor

(b) Operational reliability on a large scale - without standby units - and a proven availibility of over 90 %

(c) Clear liquor alkaline scrubbing in a closed loop employing special water-soluble additives eliminating the problems of growing plugging or scaling inherent in conventional lime-based FGD

(d) Abolishment of $CaSO_3$ sludge deposits and the associated costly fixation

(e) High-grade gypsum as a by-product derived through an integrated step in the desulfurization process

(f) Minimum operating and maintenance manpower requirements

(g) Low capital investment and economic annual overall costs from the combination of the modular gas scrubber units and the "clear liquor" process with low L/G ratios.

THE SAARBERG–HOLTER COMPANY

SAARBERGWERKE AG, with headquarters in Saarbruecken, State of Saar, West Germany, started in pit mining several hundred years ago. Today, the company has grown into a gigantic, integrated energy enterprise of about $ 1,7 billion turnover, with pit coal and its refinement still of central importance.

SAARBERGWERKE AG is the second largest West German coal producer and, with its power plants based on pit coal, is one of the most important sources of energy in the western European power network. Its generating stations at various locations account for a 50 % share of the total State of Saar power production.

SAARBERGWERKE's largest shareholders are the Federal Republic of Germany with 74 % participation, and the State of Saar with a 26 % share.

On the international scene, SAARBERGWERKE AG has technology exchange agreements with the British National Coal Board which, in turn, has similar agreements with the United States, Canada, and Australia. This worldwide exchange of technological know-how covers the areas of mining developments, coal gasification and liquification, fluidized bed combustion, desulfurization of stack gas, and chemical activities involving coal.

SAARBERG-HOLTER UMWELTTECHNIK GmbH (SHU), is a joint venture between SAARBERGWERKE and HOLTER KG, an engineering firm engaged in pollution control for underground mining equipment and foundries.
SHU's principle purpose is the design, erection, and operation of turnkey, second-generation flue-gas desulfurization systems for power plants, industry and Clans-units. Moreover SHU engineered and constructed gas cleaning plants in cokeries for tar and particulate removal.

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OPERATIONAL EXPERIENCE WITH THREE 20 MW PROTOTYPE FLUE GAS DESULFURIZATION PROCESSES AT GULF POWER COMPANY'S SCHOLZ ELECTRIC GENERATING STATION

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Southern Company Services, Inc.
Birmingham, Alabama

ABSTRACT

The two 40 MW (nominal) Babcock & Wilcox pulverized coal fired boilers at the Scholz Electric Generating Station of Gulf Power Company were retrofitted with three prototype flue gas desulfurization processes. These processes are:

1. A concentrated-mode, lime-regeneration sodium/calcium dual alkali process supplied by Combustion Equipment Associates, Inc./Arthur D. Little, Inc.
2. The CHIYODA THOROUGHBRED 101 (CT-101) process, supplied by Chiyoda International Corporation, a subsidiary of Chiyoda Chemical Engineering and Construction Company, Ltd.
3. The Foster Wheeler Energy Corporation/Bergbau-Forschung GmbH dry adsorption process supplied by Foster Wheeler Energy Corporation.

This paper presents a description of the systems' performance during an evaluation program that was conducted during 1975 and 1976. It is a summary of a complete report that is being published by The Electric Power Research Institute during the last quarter of 1977.

ACKNOWLEDGMENTS

The authors would like to acknowledge several individuals for their invaluable assistance to the evaluation program.

The value of the support and understanding of William B. Harrison, Senior Vice President of Southern Company Services, Inc. (SCS) cannot be overstated. We also wish to acknowledge George O. Layman, the Director of Power Supply for Gulf Power Company (Gulf), for his cooperation and assistance throughout the program. We greatly appreciate the efforts of William T. Lyford, Scholz Plant Manager, Ralph Calhoun, Operations Superintendent at Scholz, and James Kelly, Scholz Plant Engineer, for their support and understanding throughout the program. Without the continuing support of these individuals, the program could not have moved forward.

In addition, we wish to acknowledge the cooperation of Robert F. Ellis, Jr., President of Gulf Power Company, and Edward L. Addison, Executive Vice-President of Gulf Power Company, individually and as representatives of their organizations, in making the Scholz site available for the evaluation program.

John Craig, former Senior Research Engineer of SCS provided invaluable leadership and direction to the program during its initial stages. In addition, several other individuals at SCS and Gulf contributed invaluable support to the program:

SCS Contributing Staff

Roy Clarkson
Steve Ellis
David Morris
Tim Newton
Patsy Saint
Jane Turner

Gulf Power Contributing Staff

Jim Barbee
B. E. Charles
Hewitt Courington
Ernie Dixon
Arnold Dodson
Joe Hargrove
M. D. Hayes
Roland Howell
Roy Johns
James Knipper
J. B. Mears
John Miller
Billy Pate
Rufus Sauls
P. V. Shelfer

The cooperation, support and contributions of Arthur D. Little, Inc. (ADL), Combustion Equipment Associates, Inc. (CEA), Chiyoda International Corporation (CIC) and Foster Wheeler Energy Corporation (FW) personnel were invaluable to the completion of the evaluation program. Although several individuals from these organizations made significant contributions, the authors would specifically like to acknowledge the support of:

William Bischoff (FW)
Thomas Frank (CEA)
Charles LaMantia (ADL)
Richard Lunt (ADL)
Massaki Noguchi (CIC)

In addition, several individuals provided valuable support to the program through their participation in a project review committee. They are:

Stuart Dalton (EPRI)
Norman Kaplan (U.S. EPA)
Archie Slack (SAS Corporation)
Kurt Yeager (EPRI)

OPERATIONAL EXPERIENCE WITH THREE 20 MW PROTOTYPE FLUE GAS DESULFURIZATION PROCESSES AT GULF POWER COMPANY's SCHOLZ ELECTRIC GENERATING STATION

1.1 Project History

In 1972, the Southern electric system*, recognizing the need to develop strategies for compliance with federal emission regulations for coal-fired electric generating stations, initiated programs under which both precombustion cleaning of fuels and postcombustion gas cleaning could be furthered. This report is concerned with the postcombustion cleaning flue gas desulfurization (FGD) evaluation program conducted by Southern Company Services, Inc. (SCS), in conjunction with Gulf Power Company, for the Southern electric system. Since the program's inception in 1972, the Southern system has invested over $13.5 million. Additional investments by The Electric Power Research Institute (EPRI), The U.S. Environmental Protection Agency (EPA) and the process suppliers combine to bring the total investment in the program to over $18 million.

The processes studied were (1) a concentrated-mode, lime-regeneration sodium/calcium, dual alkali process supplied by Combustion Equipment Associates, Inc./Arthur D. Little, Inc. (CEA/ADL); (2) the CHIYODA THOROUGHBRED 101 process (CT-101)**, based on the absorption of sulfur dioxide in dilute sulfuric acid, developed by Chiyoda Chemical Engineering and Construction Company, Ltd. and supplied in the United States by Chiyoda International Corporation (CIC); and (3) the Foster Wheeler Energy Corporation/Bergbau-Forschung GmbH (FW/BF) dry adsorption process, based on the adsorption of sulfur dioxide in a bed of activated carbon, called char.

These three prototype systems (approximately 20 megawatts equivalent each) are located at the Scholz electric generating station of Gulf Power Company near Chattahoochee, Florida. The two 40 megawatt (nominal) Babcock & Wilcox, pulverized coal fired, dry bottom boilers there are fitted with Buell cold side electrostatic precipitators designed for 99.5% particulate removal. These precipitators are designed to provide selective compartmental deenergization to allow variation of particulate loadings to the FGD system for experimental purposes. Figure 1.1-1 represents a scale plan view of the entire plant property, including the experimental FGD processes. Figure 1.1-2 is a close-up scale plan view of the processes.

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* The Southern electric system is an electric utility holding company operating in the Southeast. It includes Alabama Power Company, Georgia Power Company, Gulf Power Company, Mississippi Power Company, and Southern Company Services, Inc.

** The term CHIYODA THOROUGHBRED 101 is a registered trademark of Chiyoda Chemical Engineering and Construction Company, Ltd.

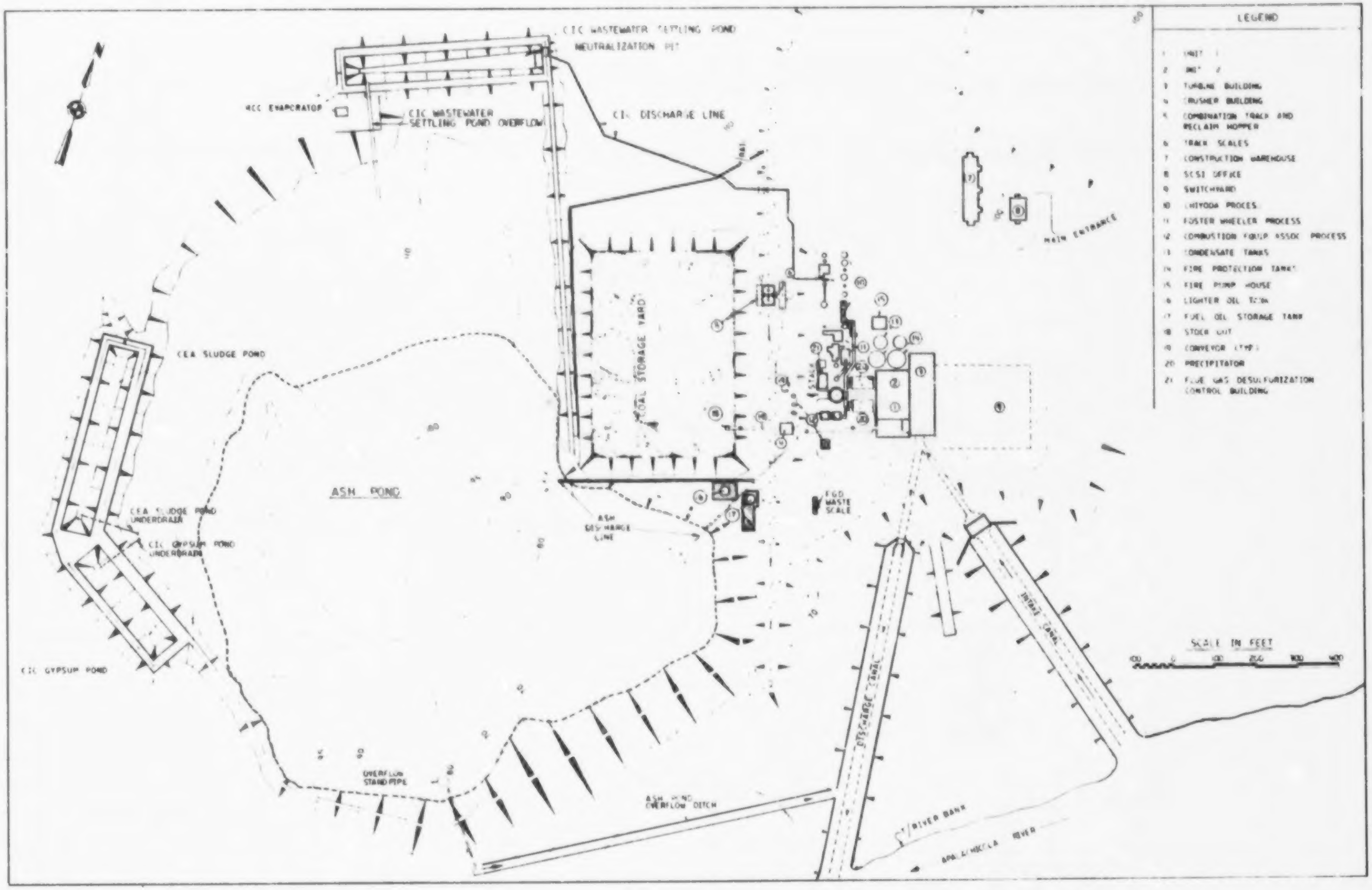

FIG. 1.1-1. Overview of the Scholz Electric Generating Station

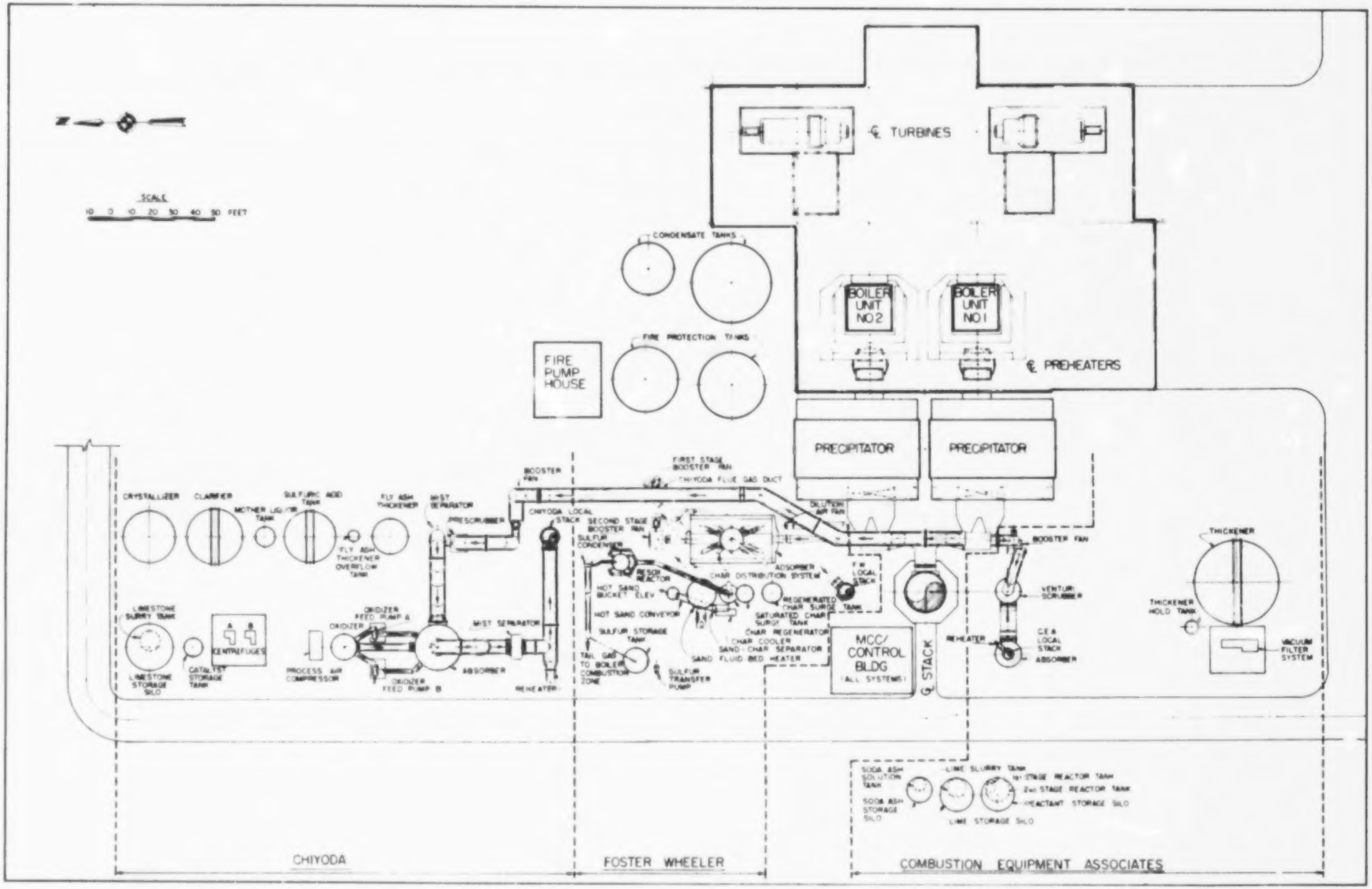

FIG. 1.1-2. Experimental Flue Gas Desulfurization Processes
Scholz Electric Generating Station

1.2 Program Scope and Objectives

The objective of the prototype test program was to establish the applicability of each prototype process for control of sulfur dioxide emissions from coal-fired steam generators. To achieve this objective, the processes were evaluated against the following criteria:

A. Reliability/Operability Potential

B. Sulfur Dioxide Removal Capability

C. Nitrogen Oxides Removal Capability

D. Particulate Removal Capability and the Impact of Particulates on System Operation

E. Impact of Contaminants on System Operation

F. Secondary Environmental Impact

G. Operating Costs

1.3 CEA/ADL Dual Alkali Process

1.3.1 System Design - In the CEA/ADL dual alkali process, Figure 1.3-1, flue gas passes through an absorption section, Figure 1.3-2, (at Scholz a variable-throat venturi and tray tower, 2 trays, in series) where sulfur dioxide, chlorides, and sulfur trioxide are removed via contact with a solution of sodium salts. The sodium/sulfur salts produced are reacted with hydrated lime in a special two-stage reactor system, Figure 1.3-3, to regenerate the sodium to an active form. The calcium/sulfur solids produced in the reactor system are separated from the liquor containing regenerated sodium compounds (via settling and filtration, Figure 1.3-4) and, at Scholz, trucked to a disposal pond. The regenerated liquor is recirculated to the absorption section for further sulfur dioxide removal.

The system was designed with the flexibility of operating in a direct lime or limestone mode as well as several dual alkali modes. The venturi was included for testing simultaneous particulate and sulfur dioxide removal. Furthermore, modifications to the system following startup provided for operation of the venturi alone by bypassing regenerated liquor around the tray tower.

1.3.2 Process Chemistry - In the absorption section of the concentrated-mode, sodium-based dual alkali process, absorption of sulfur dioxide in sodium sulfite solutions occurs to produce a bisulfite scrubber effluent solution according to the overall reaction*:

$$Na_2SO_3 + SO_2 + H_2O \longrightarrow 2NaHSO_3 \qquad (1)$$

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* For a detailed discussion of dual alkali chemistry, see Reference 1.

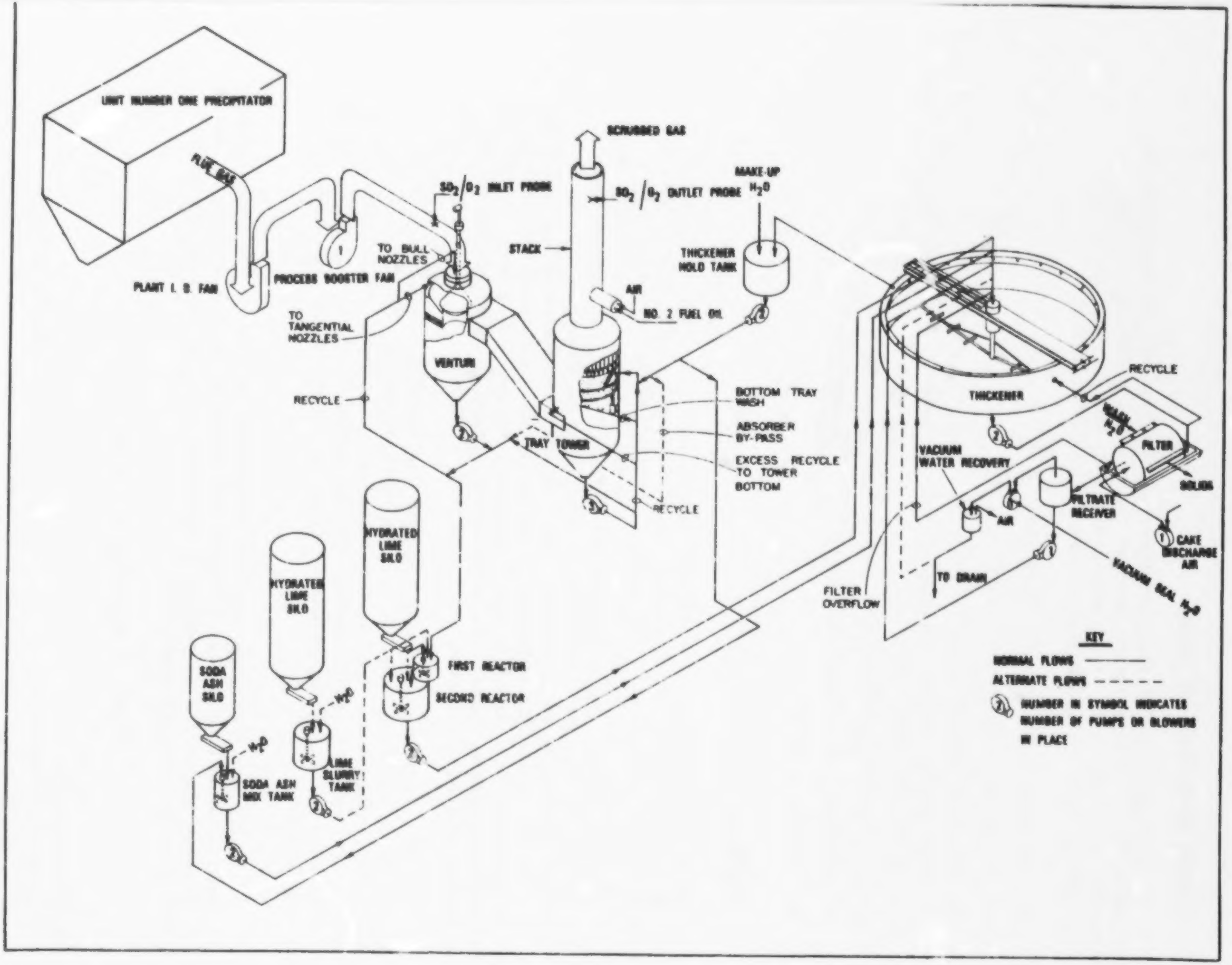


FIG. 1.3-1. CEA/ADL Dual Alkali Process Flow Diagram - Scholz Station

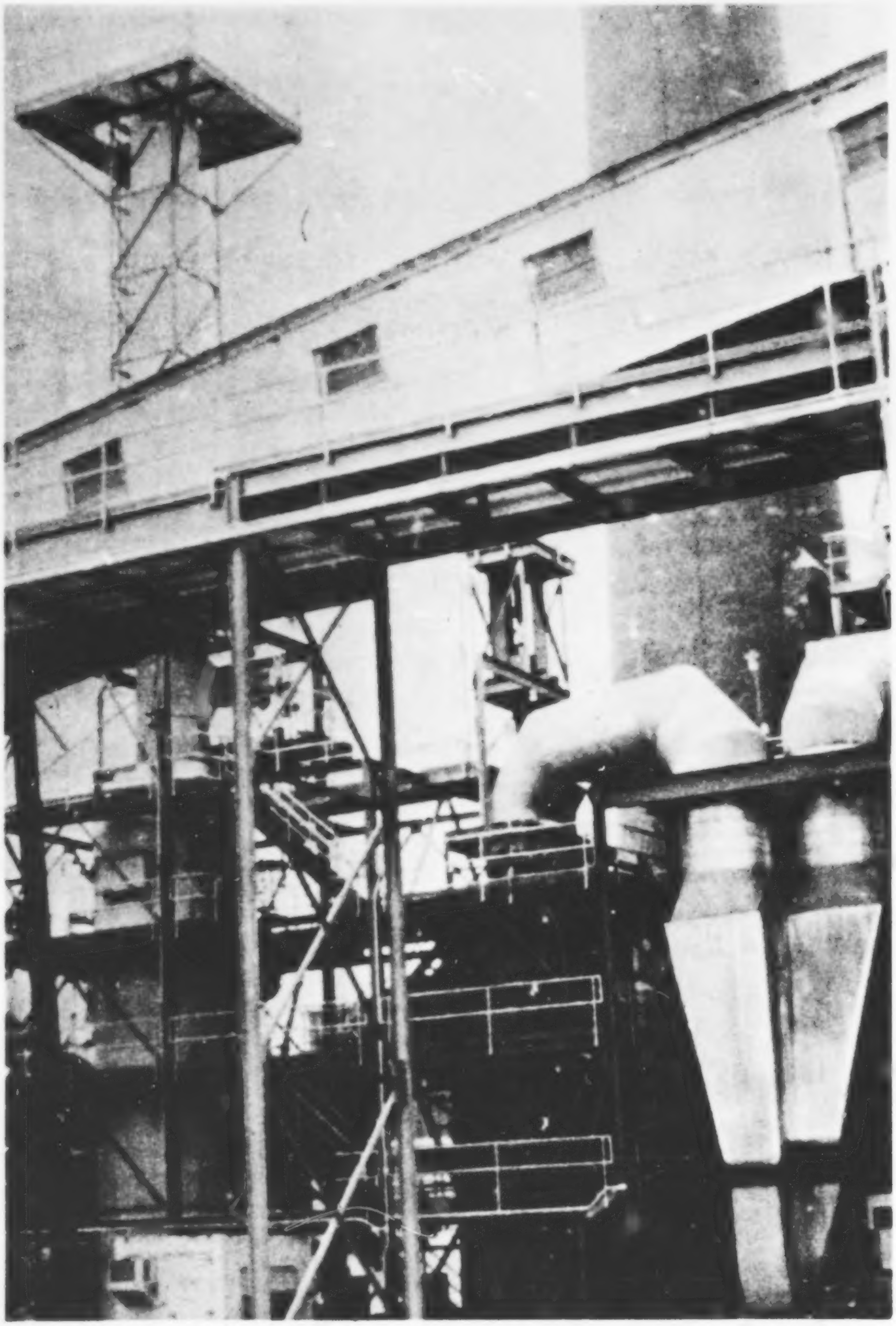
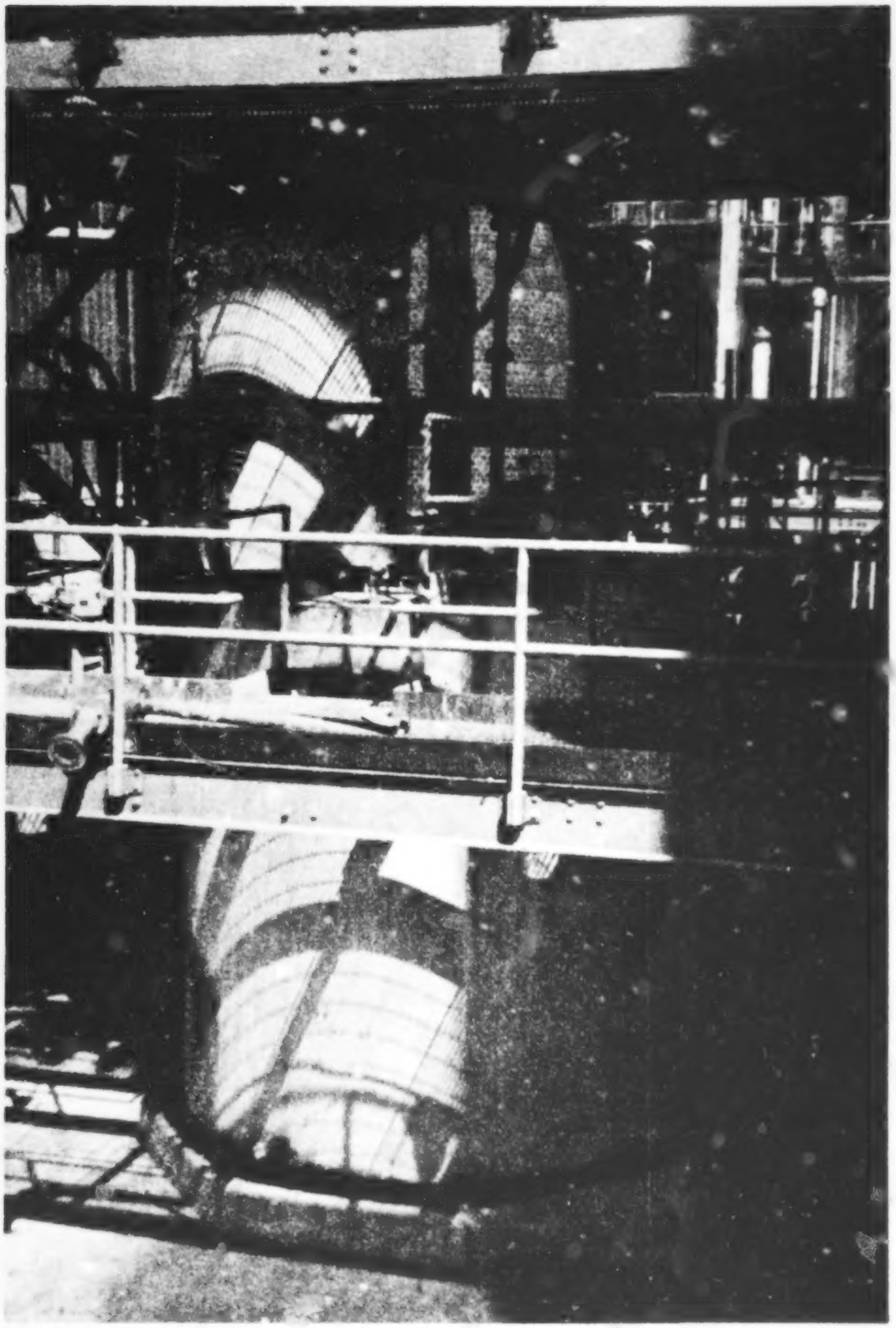


FIG. 1.3-2. CEA/ADL Dual Alkali Process Absorption Section

FIRST STAGE REACTOR



SECOND STAGE REACTOR

FIG. 1.3-3. CEA/ADL Dual Alkali Process Reactor System

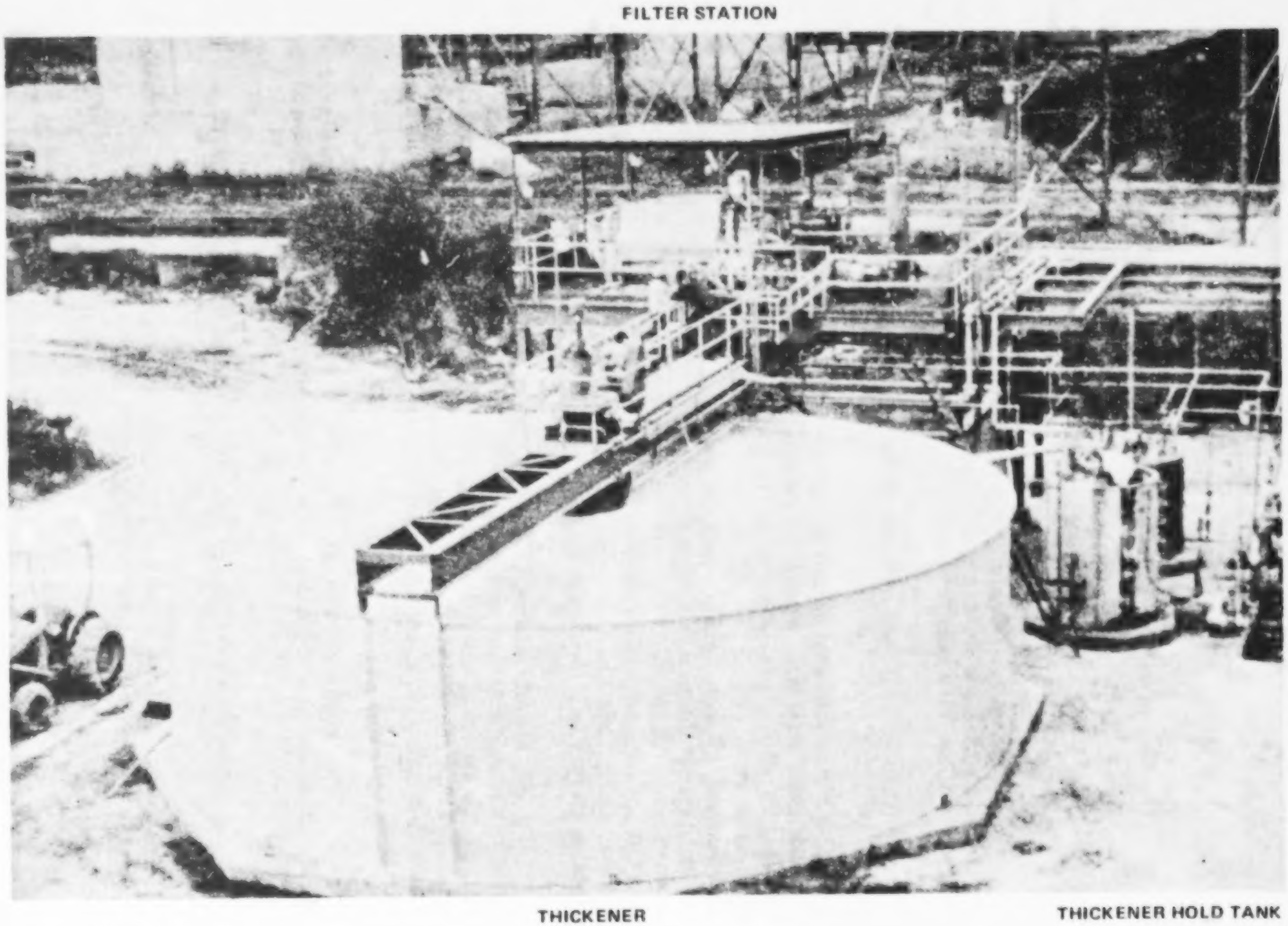


FIG. 1.3-4. CEA/ADL Dual Alkali Process Solids Processing System

The feed to the absorber will also contain sodium carbonate (used as sodium makeup to the system) and normally, depending on the extent of regeneration of absorbent, will include sodium hydroxide. Both sodium carbonate and hydroxide form sodium sulfite on absorption of sulfur dioxide:

$$Na_2CO_3 + SO_2 \longrightarrow Na_2SO_3 + CO_2\uparrow \qquad (2)$$

$$2NaOH + SO_2 \longrightarrow Na_2SO_3 + H_2O \qquad (3)$$

The sodium sulfite produced reacts further, as in equation (1), to produce more bisulfite.

During absorption, and to a lesser extent throughout the remainder of the system, some sulfite is oxidized to sulfate, converting an active form of sodium to an inactive form:

$$Na_2SO_3 + \tfrac{1}{2}O_2 \longrightarrow Na_2SO_4 \qquad (4)$$

The rate of oxidation or oxygen transfer in the absorber is a function of the absorber design, oxygen concentration, temperature of the flue gas, and the nature and concentration of the species in the scrubbing solution. The oxidation rate, in moles of sulfate produced per unit time, is relatively independent of the sulfur dioxide removal rate and is essentially constant for a given scrubber configuration and set of flue gas and process liquor conditions. This implies that the oxidation of sulfite in the scrubber is limited by oxygen mass transfer[1]. Thus, as the amount of gas/liquid contacting or the level of oxygen in the flue gas increases, the rate of sulfite oxidation (moles/unit time) will correspondingly increase.

At steady state, the sulfate produced must leave the system either as calcium sulfate, as sodium sulfate, or as a mixture of the two at the rate that it is being formed in the system. As a consequence, in order to avoid unacceptable sodium sulfate purge requirements, modes of sodium regeneration capable of precipitating calcium sulfate as well as calcium sulfite are required.

The regeneration of the sodium sulfite/sulfate effluent solutions can be considered a sequential reaction first involving neutralization of the bisulfite:

$$2NaHSO_3 + Ca(OH)_2 \longrightarrow Na_2SO_3 + CaSO_3\cdot\tfrac{1}{2}H_2O + \tfrac{3}{2}H_2O \qquad (5)$$

And then carried beyond neutralization to produce caustic, if excess lime is added, to some equilibrium hydroxide concentration:

$$Na_2SO_3 + Ca(OH)_2 + \tfrac{1}{2}H_2O \longrightarrow 2NaOH + CaSO_3\cdot\tfrac{1}{2}H_2O \qquad (6)$$

The system was designed to operate in the concentrated active sodium mode (active Na^+ concentration greater than 0.15M). In this mode, sulfate removal cannot be accomplished by the precipitation of gypsum since the high sulfite levels prevent the soluble calcium concentration from reaching a level required to exceed the gypsum solubility product[1]. However, calcium sulfate is precipitated along with calcium sulfite in the regeneration reactor, resulting in a solid solution of the two salts. Depending upon the concentration of sulfite and sulfate and the pH of the solution, the following reaction for sulfate removal also occurs simultaneously with reactions (5) and (6):

$$Ca^{++} + SO_4^{=} + \tfrac{1}{2}H_2O \longrightarrow CaSO_4 \cdot \tfrac{1}{2}H_2O \qquad (7)$$

After regeneration, the solids are separated from the regenerated liquor and washed. If the level of sulfate formation is matched with the level of sulfate precipitation, all the sulfate formed in the system can leave as calcium salt. In that case, no soluble sulfate purge is necessary to maintain a sulfate balance. In practice, even if such a balance is established, the washed calcium sulfite/sulfate salts will contain some soluble sodium salts as well as soluble flue gas constituents which must be purged, and some sodium makeup to the system will therefore be required.

In principle, any sodium lost with the washed waste solids can be replaced by the addition of sodium hydroxide, since carbonate softening is not required in concentrated-mode dual alkali systems, or by sodium sulfate if oxidation rates are low enough. Neither of these options were tested at Scholz. Sodium losses there were balanced by the addition of sodium carbonate (soda ash).

1.3.3 System Operation - The system became operational on February 3, 1975, and was operated over a period of 17 months through July 3, 1976, when it was shut down at the completion of the test program. During these 17 months, it logged over 7100 hours of operation with shutdown times of varying length for maintenance, modifications, repairs, and when the boiler was shut down. The 17 months of operation can be broken down into three discrete operating periods, Table 1.3-1, defined by flue gas composition, coal characteristics, and system operation.

During the first operating period, covering the first five and one-half months of operation, the system treated flue gas with significantly lower sulfur dioxide concentrations and higher oxygen levels than the range for which it was designed. Sulfur dioxide concentrations in the flue gas averaged about 1050 ppmv compared with the minimum design concentrations of 1800 ppmv. Sulfur dioxide removal averaged 93%. The actual inlet concentrations varied over a range from 600 to 1550 ppmv, often fluctuating daily and even hourly over the entire range of concentrations. Furthermore, oxygen concentrations in the flue gas entering the scrubber ranged from 5 to 11% by volume (equivalent to 30 to 100+% excess air) as compared to a maximum design concentration of 6.5%. Operation with these high levels of oxygen in the flue gas in combination with a relatively low sulfur dioxide concentration represented a difficult task for the system, especially at lower load conditions. (See Section 1.3.4.1.1 below for a discussion of these effects.)

TABLE 1.3-1

CEA/ADL DUAL ALKALI PROCESS
SUMMARY OF OPERATING PERIODS

PERIOD	INTERVAL	OPERATING HOURS	COAL SULFUR	TESTING
1	2/3/75-7/18/75	2537	Low	Startup & Shakedown
2	9/16/75-1/2/76	2180	Low/Medium	Stable Load
3	3/16/76-7/3/76	2411	High	Fluctuating Load Particulate

A slightly higher sulfur coal delivered in August of 1975 resulted in marginally higher sulfur dioxide concentrations during the second operating period. During this period, which spanned three and one-half months, inlet sulfur dioxide concentrations ranged from 800 to 1700 ppmv and averaged about 1200 ppmv. Sulfur dioxide removal averaged 95%. Oxygen levels in the flue gas were reduced from the previous 5 to 11% range to the 5 to 7% range in September after air preheater repairs and modifications to boiler operating conditions. These slight improvements in the oxidation potential of the system, along with the philosophy of operating the process at slightly higher dissolved solids levels than in period 1, combined to reduce oxidation and to permit better control over system operations, including reduction in the sodium losses (and sodium sulfate purge) in the filter cake.

In February, 1976, prior to the third operating period, which continued for four months, a higher sulfur coal, containing an average of 3.5 wt% sulfur, was delivered. This higher sulfur coal allowed improved system operation and testing on inlet sulfur dioxide concentrations ranging between 1500 and 2850 ppmv. Sulfur dioxide removal averaged 95%. In addition to stable load testing as conducted during periods 1 and 2, testing during period 3 also included fluctuating gas load and particulate testing. During the particulate tests, the electrostatic precipitator was partially or completely deenergized.

1.3.4 System Performance

1.3.4.1 Reliability/Operability Potential - The purpose of this installation was to test the process chemistry and design on a relatively small scale in order to evaluate the viability of the technology. While reliability was a principal concern, the system was not intended to be a demonstration unit nor a test of the ultimate reliability of such systems when applied full-scale. Every attempt was made to operate the dual alkali system at Scholz in a continuous manner, as if it were a full-scale production facility. However, it was the first concentrated mode, lime-regeneration sodium/calcium dual alkali system designed for application on a utility boiler to be constructed in the U.S. on larger than a pilot plant scale. As a result, several design inadequacies were identified which one would not expect to see repeated in future systems. Consequently, it is most accurately described as a "prototype" system.

The design was based upon scale-up by a factor of about 40 from the CEA/ADL dual alkali pilot plant in Cambridge, Massachusetts. Although the system contained spare pumps throughout, other key elements such as the reactor system, lime feed system, and filter station were not spared in this application. Multiple installations and/or spare capacity for these critical items will normally be incorporated in full-scale applications. In spite of these considerations and the difficult flue gas conditions encountered during the first operating period, the operating record of this prototype unit over the 17 months of operation from initial startup through the completion of the test program is quite impressive.

In an effort to put the operation of the process on a clearly visible basis, the 4 PEDCo[2] FGD system viability parameters have been calculated for the system. These parameters are presented as cumulative percentages in Table 1.3-2. In addition, the operability parameter on a monthly basis is presented in Figure 1.3-5.

Unfortunately, none of these parameters is adequate for evaluating the future performance potential of first-of-a-kind prototype systems. Even for full-scale systems, they cannot be used literally to judge an entire technology by results at only one site. They are, however, the only quantitative measures of FGD system operation that approach universal acceptance and for that reason have been presented here. What is needed for a proper evaluation is long-term operation on several large, base-loaded boilers. Lacking that, a qualitative judgment of the system's potential for acceptable long-term operation is the only means of evaluation available.

1.3.4.1.1 Process Performance - The process operability (ease of operation) was excellent in all respects. The system was successfully operated over a range of widely fluctuating inlet flue gas sulfur dioxide levels (1.6 to 4 wt% sulfur fuel), oxygen concentrations, and flow rates with little or no change in the sulfur dioxide removal efficiency, waste cake properties, or lime utilization. It demonstrated an ability for continuous operation with large and frequent variations in pH in both the scrubber circuit (4 to 7) and the regeneration/solids dewatering section (6.5 to 13).

The most impressive aspect of system operation was its resistance to short-term upsets. Low soluble calcium levels throughout the system, even during most upset conditions, resulted in a low potential for calcium/sulfur salt precipitation, particularly in the scrubber circuit. In addition, operation for over 4500 hours without mist eliminator washing and without any mist eliminator scaling confirmed the viability of system operation without a mist eliminator wash.

The dual alkali process does have a limitation in that there is a minimum level of sulfur content in the fuel below which it cannot be successfully operated in the concentrated mode. This lower limit is a function of the rate of oxidation in the system and is therefore dependent not only on the sulfur content in the fuel, but also the level of oxygen in the flue gas. For typical pulverized-coal-fired boilers, the design of a concentrated-mode dual alkali system for use on flue gases produced from the combustion of coal containing between 1 and 2 wt% sulfur, will have to be considered carefully on a case-by-case basis. For fuels containing less than 1 wt% sulfur, a concentrated-mode dual alkali system cannot be operated at the excess air levels that are typical for pulverized-coal-fired boilers without an intentional purge of sodium sulfate, either in the filter cake or as a separate purge stream. However, above 2 wt% sulfur in the fuel, and in many cases for coal containing between 1 and 2 wt% sulfur, the operation of the system is excellent. In fact, overall system operation improves significantly as higher sulfur content fuel is utilized.

TABLE 1.3-2

CEA/ADL DUAL ALKALI PROCESS
VIABILITY PARAMETERS
(FEBRUARY 1975-JULY 1976)

PARAMETER	VALUE (%)	TOTAL TIME NUMERATOR (HOURS)	TOTAL TIME DENOMINATOR (HOURS)
Availability	78	9679	12,376
Reliability	80	7128	8911
Operability	70	7128	10,172
Utilization Factor	58	7128	12,376

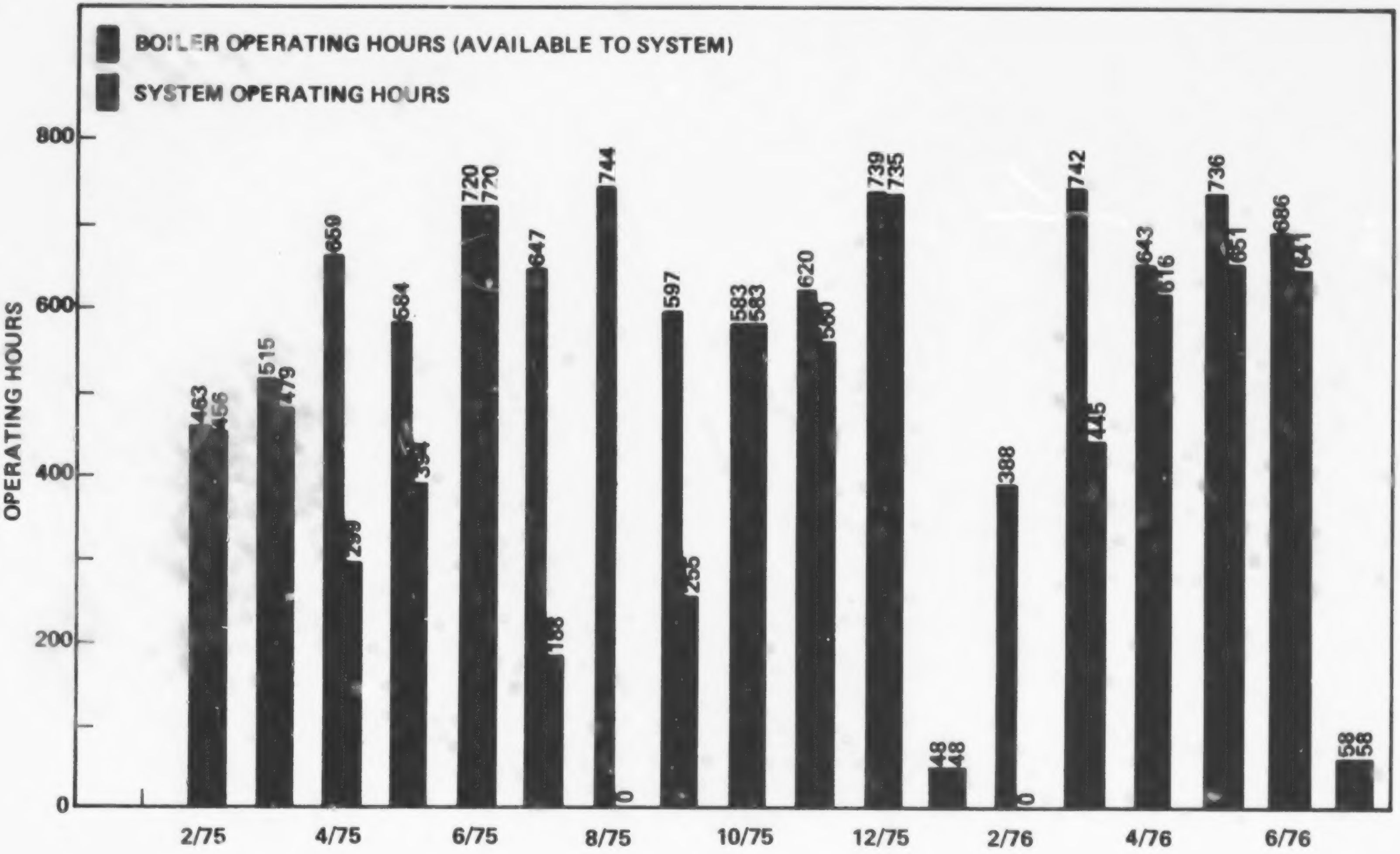


FIG. 1.3-5. Operability of the CEA/ADL Dual Alkali Process

It is difficult to generalize with respect to the long-term reliability potential of any mechanical/chemical process. However, the results at Scholz indicate that the overall performance of a properly designed and operated dual alkali system should be superior to that of direct lime and limestone systems because;

- As discussed above, the system is highly resistant to upset. The potential for calcium/sulfur salt precipitation (scaling) is eliminated by the high sulfite levels in solution, except in extreme upset conditions.

- The handling of slurries in the absorption section is completely eliminated.

- The most important control parameter pH, has a wide, acceptable range of operation and, unlike on lime and limestone systems, has no effect on scaling up to values of 12.5 to 13 in the reactor system.

1.3.4.1.2 Equipment Performance - Overall, the mechanical performance of the system was quite good. Most of the problems encountered with equipment and instrumentation during the course of the test program were mechanical in nature and reflected design or fabrication oversights commonly associated with a prototype system. All but a few, as discussed below, were resolved during the course of the test program by simple operational adjustments and/or equipment modifications.

A. Mechanical Problems - Aside from occasional failures of rubber-lined pumps, the most significant equipment problems encountered in the system involved the filter, vessel liners, liquid control and block valves, and solids buildup in the first-stage reactor. Collectively, these accounted for the bulk of mechanical-related downtime and maintenance.

(1) The filter (rotary drum, vacuum type) was the greatest source of trouble in the system. Normally, filters do not require an inordinate amount of maintenance. However, to avoid corrosion problems anticipated with low pH and high chloride levels (if the system had been operated in the limestone dual alkali or direct limestone modes), the filter was fabricated from plastic and fiberglass. Unfortunately, there were frequent failures at stress points as well as erosion of the plastic and fiberglass parts. The filter was returned to the manufacturer between operating periods 2 and 3 for modification and overhaul to improve the operation. After its return, maintenance requirements were significantly reduced, but were still excessive. The use of stainless steel for the construction of this equipment will avoid most of the problems experienced at Scholz.

(2) The test stack and venturi throat were lined with an acid-resistant cement (Sauereisen 33). While the cement liner in the venturi throat deteriorated to some extent over the course of the program, its failure rate was low enough to be within the realm of normal maintenance. After patching a few soft spots and cracks from the original installation, the cement liner in the stack gave excellent service.

The liner system used in the process vessels was of the flake-filled polyester type (Heil 4850). Several failures of this liner occurred during the course of the testing. The most serious were related to (a) poor quality control during application, (b) abrasion, and (c) excessive temperature and/or temperature cycling. Quality control related failures occurred as a few random pinholes and as failure of large sections of liner in the thickener and thickener hold tank. The pinhole failures required only a few hours to repair. The large-scale failures required several weeks to repair in the interim between periods 2 and 3. After repairs, these liners performed acceptably. Abrasion-related failures occurred on the second-stage reactor floor under the agitator and at the tangential nozzle discharge point above the throat in the venturi. The liner in these areas was replaced with an abrasion-resistant formulation (Heil 413G) which gave acceptable service over the remaining 2400 hours of testing. The temperature-related failures occurred on an extension of the gas inlet duct located in the gas quench zone in the venturi. Attempts to repair this failure by replacing the flake-filled polyester liner with a high-temperature vinyl ester liner were not successful. Although the polyester liner appears to have failed from excessive temperature, the exact cause of the failure of the vinyl ester is not known. It may have been related to poor quality control during application and/or temperature cycling in the quench zone. These failures suggest that corrosion-resistant metal alloys may be necessary in such areas.

(3) Erosion failures of rubber linings occurred in several block and control valves over the course of the program. The failures of the block valves were caused when they were forced into throttling service to restrict the flow from the oversized venturi and absorber pumps. (The pump oversizing was a result of the inclusion of direct lime/limestone scrubbing capability into the system). The problems with block valves were resolved by slowing the impellers and decreasing the impeller diameter in some of the pumps and/or installing orifices to decrease the pump output.

The failures of the control valves were related to excessive throttling (caused by the aforementioned pump overcapacity) and/or "tramp" material in the system. In any event, the rubber-lined control valves were replaced by 316SS valves. These stainless valves showed no signs of deterioration after 2400 hours in the absorption section (pH 4 to 6). Other stainless control valves were originally installed in high pH service (pH 8 to 13) in the regeneration area. These, too, were in excellent condition after 7100 hours.

(4) Buildup of product solids (calcium sulfite and sulfate) occurred in the first-stage reactor throughout the test program. Through adjustments made to the reactor system and simulation of the operation in the CEA/ADL pilot plant, the cause of the problem was traced to poor agitation and inadequate lime feed control during severe upset conditions (e.g., gross overfeeding of lime). While the buildup was never serious enough to cause a shutdown, it did require occasional cleaning. Improved agitation and better process control should reduce such cleanings to, at worst, two to three times per year on future systems. Such maintenance would not normally require a system shutdown in large-scale systems where parallel reactor trains can be used, or the first-stage reactor can temporarily be bypassed.

B. Instrumentation Problems - The process instrumentation performed well, overall, Instrumentation problems that affected system operation involved pH measurement and liquid level transmitters.

(1) The flow-through pH probes originally installed in the system were prone to plugging and/or erosion and failure of the probe tips. The flow-through unit in the reactor system was replaced with an immersion unit which proved to be much more reliable. Modification of the liquid take-off lines to the flow-through units in the absorber area, to provide minimum flow restriction, and increasing the flow rate minimized problems with these units.

(2) The liquid level transmitters installed originally were unreliable and required an inordinate amount of instrument maintenance. They were eventually replaced with Foxboro units, which proved to be much more reliable and less prone to failure of critical parts.

1.3.4.2 Sulfur Dioxide Removal Capability - The system was operated using two different configurations for sulfur dioxide removal: the venturi and absorber together in series (with two trays) and the venturi alone, in an attempt to limit oxidation during operation on lower sulfur coals. In the latter configuration, the trays were not removed from the absorber; rather, the regenerated liquor normally fed to the top tray was diverted either to the absorber recycle tank (from there it was transferred to the venturi) or bypassed directly to the venturi through a line installed in May, 1975. (The bypass line was not in the original design, since the operation on low-sulfur coal was not anticipated.) When the absorber was not used, the recycle flow to the top tray and the flow to the spray on the underside of the bottom tray were both discontinued; however, the absorber pumps were maintained in operation to transfer liquor collected in the absorber tank to the venturi. In order not to exert excessive back-pressure on the absorber pumps, a recycle was maintained through an open spray header during intervals when there was no feed to the trays.

During period 1, both operational configurations were used at different times. However, during periods 2 and 3, only the combined venturi and absorber configuration was used. The sulfur dioxide removal efficiency achieved with each of these configurations, as a function of pH, is shown in Figure 1.3-6 for intervals when the inlet sulfur dioxide ranged from 1050 to 1250 ppmv (periods 1 and 2). Figure 1.3-7 shows the sulfur dioxide removal achieved as a function of pH at the higher inlet sulfur dioxide levels (1900 to 2200 ppmv) in period 3.

The data in Figures 1.3-6 and -7, which reflect the general operating experience at Scholz, confirm the high sulfur dioxide removal capability of sodium solution scrubbing systems operating in an equivalent concentration range of active sodium (0.2 to 0.4 M Na^+). Achieving a given outlet sulfur dioxide level (within the limit of the number of contact stages in use) is essentially a matter of adjusting the operating pH of the scrubber system by changing the feed forward rate and/or pH of the regenerated liquor. Over the 15 months of operation between April, 1975 and July, 1976, the average sulfur dioxide removal, using both the venturi and absorber, was 95.5% (with low-

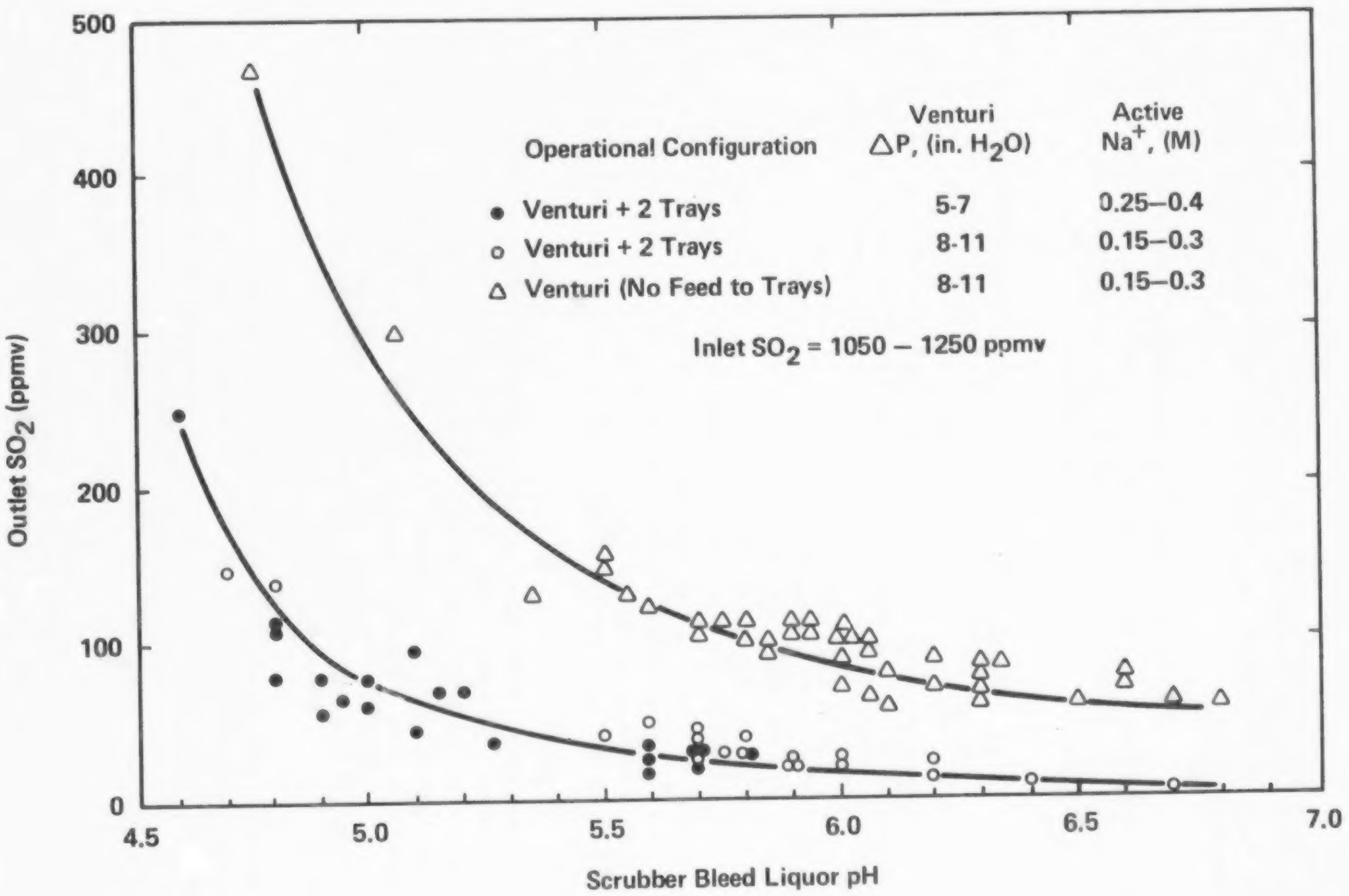


FIG. 1.3-6. CEA/ADL DUAL ALKALI PROCESS
SO_2 Removal as a Function of pH -
Low/Medium-Sulfur Coal

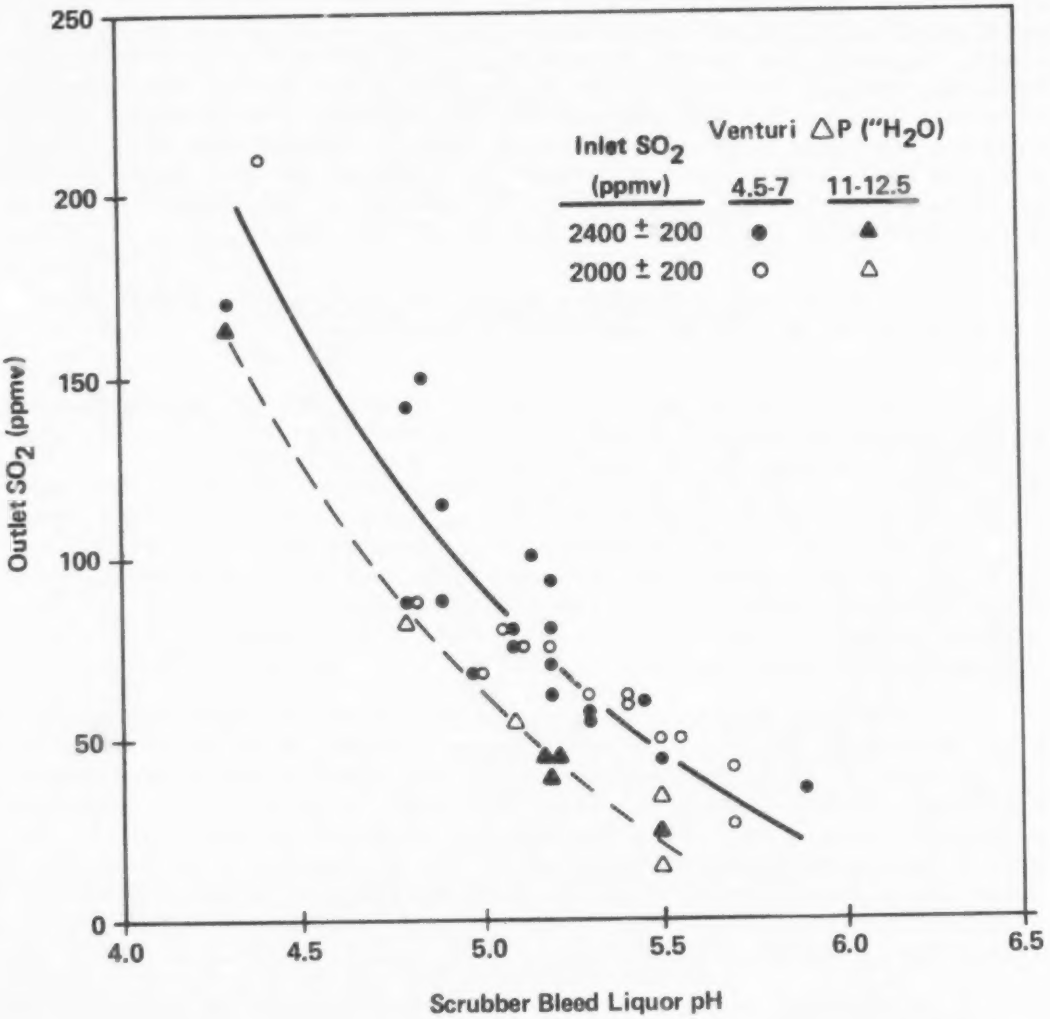


FIG. 1.3-7. CEA/ADL Dual Alkali Process SO_2 Removal as a Function of pH - High-Sulfur Coal

and high-sulfur coal). With the venturi alone, it was 90.7% (low-sulfur coal only). Operation with highly fluctuating gas flows had no effect on sulfur dioxide removal.

Because of the high concentration of alkaline ions in the scrubbing liquor, the liquid flow requirements for sulfur dioxide removal in a sodium-based dual alkali system are small. At Scholz sulfur dioxide removals in excess of 95% were easily achieved with liquid-to-gas ratios (L/G) on the order of 25 gal/1000 acf in the venturi and 5-7 gal/1000 acf in the tray tower. However, the venturi is necessary only for particulate collection. In addition, most of the liquid flow in the absorption section was necessary to eliminate deadheading the pumps, which were designed for operation in a direct limesone scrubbing mode as well as dual alkali. Liquid flow requirements for a system configured without a venturi and designed for dual alkali only would be 12 gal/1000 acf or less for 90 to 95% removal of the sulfur dioxide in flue gas produced from the combustion of 3 wt% sulfur coal.

1.3.4.3 Nitrogen Oxides Removal Capability - No significant capability for nitrogen oxides removal was noted in the system.

1.3.4.4 Particulate Removal Capability and the Impact of Particulates on the System Operation - It was felt that long-term operation with high particulate loadings in the inlet flue gas was not necessary. Since several demonstration, as well as full-scale, FGD systems of similar design are operating with full particulate loading to the system, it was felt that the long-term effects of high particulate loadings would be similar from system to system. Because of this, the electrostatic precipitator was left fully energized until the last three weeks of the test program. At that time, various sections, Figure 1.3-8, were deenergized to observe the short-term effects of high inlet particulate loadings on system performance.

The particulate removal capability of the scrubber system was tested with the venturi operating at both 12 and 17 inches of water pressure drop and at three inlet particulate loadings. The results of the particulate measurements for each test condition are given in Table 1.3-3. Standard EPA procedures were used in all measurements. As would be expected, outlet particulate loadings increased slightly with increasing inlet loadings; but, there was no statistical difference between outlet particulate loadings (or particulate removal efficiencies) measured at venturi pressure drops of 12 inches and 17 inches of water.

Normally, it would be expected that the increased pressure drop would increase removal efficiency, particularly at the higher inlet loadings. The fact that the venturi was followed by an absorber containing two cross-flow sieve trays is possibly the cause of this. Averaging all data (12 and 17 inches of pressure drop), at the highest inlet loadings (about 3 gr/dscf), the particulate removal efficiency of the scrubber system with the precipitator out of service was 98.9%.

1.3.4.5 Impact of Contaminants on System Operation - The process operated for significant periods of time with chloride concentrations in excess of 13,000 ppm in the absorption and reactor sections and 8500 ppm in the

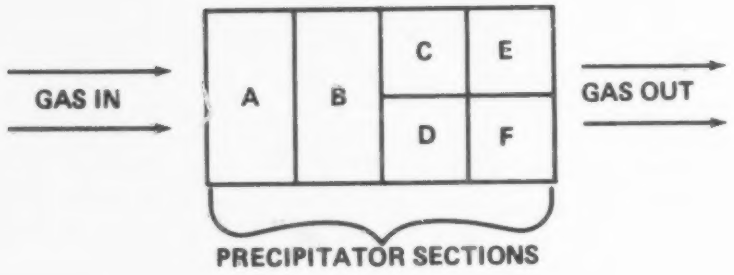


FIG. 1.3-8. Schematic of the Electrostatic Precipitator Fields at the Scholz Electric Generating Station

TABLE 1.3-3

CEA/ADL DUAL ALKALI PROCESS[1]
SUMMARY OF PARTICULATE TEST RESULTS

GENERAL CONDITIONS:

$\sim$100% GAS LOAD
REHEATER OFF
VENTURI L/G $\sim$ 20 GALS/1000 ACF$_{SAT}$

TEST NO	PRECIPITATOR SECTIONS IN SERVICE	VENTURI Δ P ("H_2O)	PARTICULATE LOADINGS INLET (grs/dscf)	OUTLET (grs/dscf)	REMOVAL EFFICIENCY (%)
1	A&B	12	0.025	(0.010)	(60)
2	A&B	12	0.017	0.012	29
3	A&B	12	0.021	0.015	29
4	A&B	17	0.018	0.011	39
5	A	17	0.084	0.026	69
6	A	12	0.034	0.021	38
7	A	12	0.047	0.027	43
8	None	12	3.60	0.037	99.0
9	None	12	3.00	0.024	99.2
10	None	12	2.95	0.034	98.8
11	None	17	2.73	0.037	98.6
12	None	17	2.29	0.033	98.6
13	None	17	3.34	0.035	99.0

[1] In interpreting these results, it should be remembered that the venturi is followed by a tray tower containing 2 cross-flow sieve trays. The tray tower contributes to the particulate removal and to some extent "masks" the effect of venturi pressure drop.

thickener/filter area without any significant effects on process operation being observed. Chlorides do affect the choice of materials of construction within the system. However, 316SS or 317 SS (or their low carbon counterparts for welded service) appear adequate for the process conditions (exclusive of reheat).

With the exception of calcium, sodium, and potassium, none of the common metal ions were found in the process/liquor in excess of 1 ppm. (Sodium and calcium are, of course, common to the system.) Two checks of potassium in the process liquor showed levels of approximately 300 and 1300 ppm. No effects were noted at these concentrations. In addition, no effect was observed from fluoride concentrations up to 70 ppm in the process liquor.

1.3.4.6 Secondary Environmental Impact - The secondary environmental impact of an FGD system is highly site specific and must be assessed on a case-by-case basis. In relation to the entrainment of particles from the system, it is a function of gas flow distribution and velocity, mist eliminator performance, and possibly condensation downstream of the mist eliminator. In relation to land use and water quality, it is dependent upon the trace constituents in the coal fired at the power plant, the local groundwater hydrology, the method of waste disposal, the local topography, and probably most importantly, the impact of the relative amounts of rainfall and evaporation at the site on the system's water balance. Depending on the combination of these parameters, the impact of an FGD system on water quality can range from practically nil, in an arid region where the soil has low permeability, to significant, in wet regions with highly permeable soils or karst limestone formations.

1.3.4.6.1 Stack Emissions - Rainout from the plume was a problem in the immediate vicinity of the dual alkali process. However, it was apparently related to (1) condensation downstream of the mist eliminator in the poorly insultated test stack, (2) the high stack velocity (about 50 fps design) which stripped liquid off the stack wall, and (3) the fact that reheat was rarely used in order to conserve fuel oil. Entrainment of process liquor to the atmosphere from the dual alkali process at Scholz was measured on two occasions. The tests, Table 1.3-4, were conducted using sodium from the process liquor as a tracer. The maximum entrainment measured during any test was 0.0035 grains of solids/dscf (0.025 grains of liquor/dscf) or 0.018 gpm.

1.3.4.6.2 Comparison With Direct Lime or Limestone Systems - The impact of a concentrated-mode sodium/calcium dual alkali system on land use and water quality will be similar to that of direct lime or limestone systems if the local conditions (as stated above) are the same. A possible exception is the concentration of total dissolved solids in the leachate or runoff from waste areas, caused by the sodium salts occluded in the waste cake. On the other hand, if a magnesium-promoted direct lime or limestone system is compared with sodium/calcium dual alkali, the total dissolved solids concentration in the liquor occluded with the waste cake will be similar, since magnesium, like sodium, is significantly more soluble than calcium.

TABLE 1.3-4

CEA/ADL DUAL ALKALI PROCESS
SUMMARY OF LIQUOR ENTRAINMENT MEASUREMENTS

DATE	SUPERFICIAL[1] GAS VELOCITY (FPS)	MIST[2] ELIMINATOR WASHED (YES/NO)	REHEATER (ON/OFF)	PRECIPITATOR[3] SECTIONS IN SERVICE	ENTRAINMENT RATE (GPM)	ENTRAINMENT RATE (GRS LIQUOR/DSCF)	ENTRAINMENT RATE (GRS SOLIDS/DSCF)
6/76	9.1	NO	OFF	ALL	0.0055	0.0076	0.0011
6/76	9.1	NO	OFF	ALL	0.0058	0.0085	0.0012
12/75	7.7	NO	ON	ALL	0.015	0.021	0.0029
12/75	7.6	NO	ON	ALL	0.011	0.015	0.0021
12/75	7.6	NO	ON	ALL	0.012	0.018	0.0025
12/75	8.1	NO	OFF	ALL	0.018	0.025	0.0035
12/75	7.9	NO	OFF	ALL	0.015	0.021	0.0029
6/76	7.2	NO	OFF	ALL	0.0039	0.0066	0.0009
6/76	7.4	NO	OFF	ALL	0.0042	0.0071	0.0010
6/76	5.0	NO	OFF	ALL	0.0028	0.0076	0.0011
6/76	5.0	NO	OFF	ALL	0.0033	0.0080	0.0011

---

[1] Design gas velocity in the scrubber was nominally 9.0 fps.

[2] Mist eliminator wash discontinued after August 1975.

[3] Inlet particulate loading less than 0.02 grs/dscf.

1.3.4.6.3 Waste Solids - The waste, consisting mainly of calcium sulfite ($CaSO_3 \cdot \frac{1}{2}H_2O$), ranged in particle size from 0.01 to 0.05 mm. The solids content of the filter cake produced throughout the test program typically ranged from 45 to 60 wt% solids. In the three operating periods described in section 1.3.3 the average cake solids content was 48, 51, and 54 wt% respectively. These data include periods when the filter operation was less than optimal, and, also, upset conditions such as contamination of the lime feed with limestone. (The limestone contamination was due to mixups in the reactant supply to the CT-101 system at the site.) While these data include upset conditions, the effect of the sulfate content in the cake (or system oxidation) is apparent. At the higher sulfate levels (produced from the increased oxidation rates for lower sulfur fuel) the solids content of the cake decreased. This confirms similar data obtained by CEA/ADL in pilot plant operations[1].

For the most part, the waste had the appearance of a moist powder or moist, fine sand and could be loaded into a conventional dump truck and hauled or conveyed by a belt conveyor to a disposal area. However, it is doubtful that the disposal area could be operated as a landfill unless the waste can be sufficiently dewatered by dry flyash addition, because without the addition of dry flyash the vacuum filtered waste exhibits low bearing strength, high settlement, and provides little traction for conventional earth moving equipment. Further study and possibly a field demonstration of landfilling CEA/ADL waste are warranted because the experience at Scholz did not determine conclusively whether or not landfilling could be successfully carried out without further dewatering the waste by dry flyash addition.

A. Disposal Methods - The CEA/ADL waste at Scholz was dumped from a truck into a narrow pond equipped with a polyethylene liner, underlain by a natural clay layer. A photograph of the pond is shown in Figure 1.3-9 and pond construction details are shown in Figure 1.3-10. This waste disposal method is somewhat of a hybrid between ponding and landfilling and was employed only for expediency due to the small size of the system and short duration of the project. A full-scale installation of the CEA/ADL process would probably employ a landfill or a pond for disposal.

(1) Landfill - For a landfill operation to be carried out successfully with CEA/ADL waste, it is probable that dry flyash will have to be mixed with the waste to obtain a solids content high enough for compaction. Laboratory and field tests indicate that sufficient quantities of dry flyash will have to be available to obtain at least 70% solids in the mixture. The dry flyash requirements for obtaining solids contents of 70 and 75% in a mixture of CEA/ADL waste and dry flyash are shown in Table 1.3-5.

Laboratory and field testing at Scholz indicated that landfilling and compacting a mixture of CEA/ADL waste and dry flyash or a mixture of CEA/ADL waste, dry flyash, and lime, could be successfully carried out. A photograph of the field compaction is shown in Figure 1.3-11. Dry densities of 75 to 85 pounds of dry waste and ash per cubic foot of fill were routinely obtained in both the laboratory and the field. The dry densities obtained in the field are listed in Table 1.3-6. The dry density obtained in landfilling is roughly twice hat obtained from ponding and would result in a disposal volume, for a given dry weight of waste and ash, of one-half the disposal volume required for ponding (See Section 1.3.4.6.3 A(2) below). Field and laboratory testing at Scholz

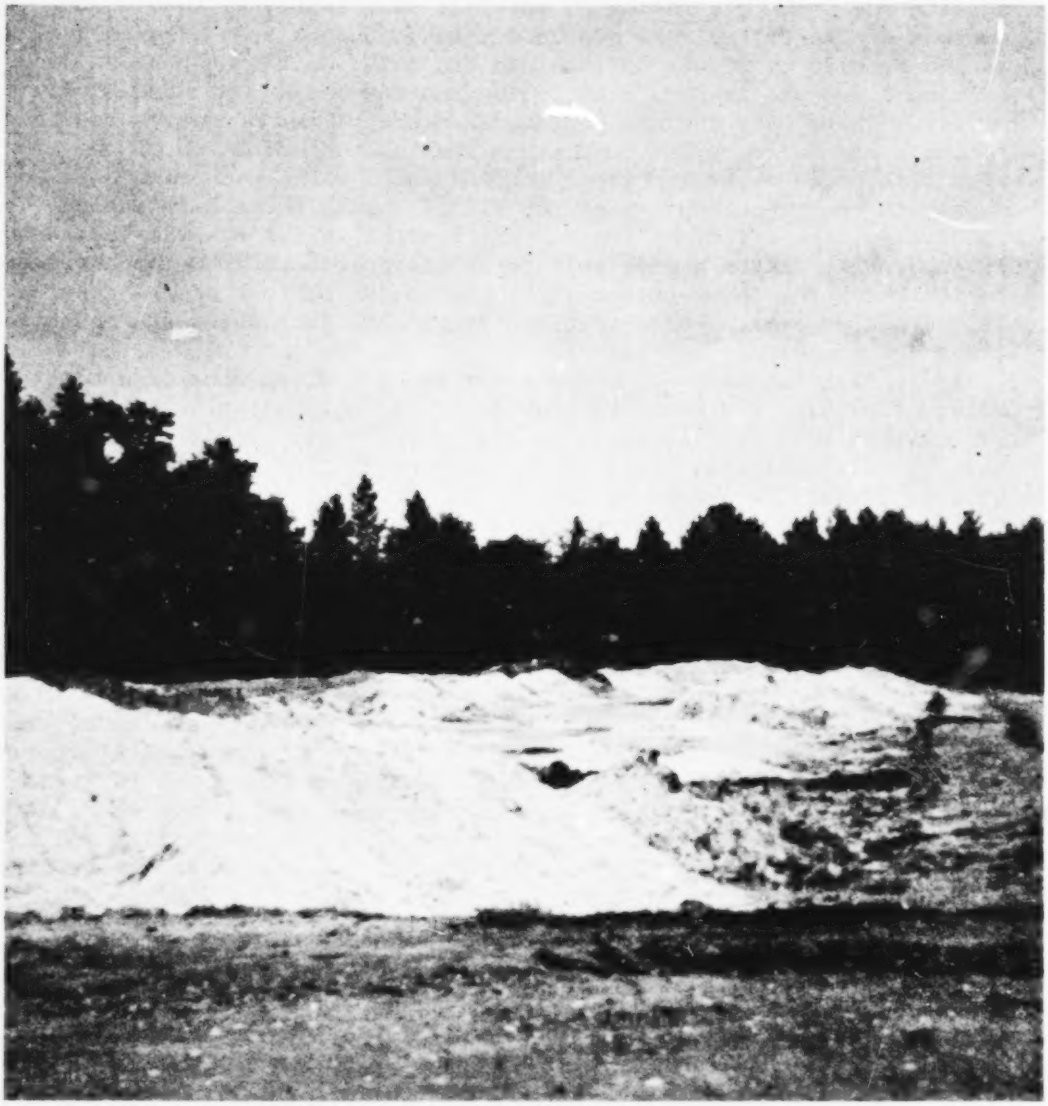
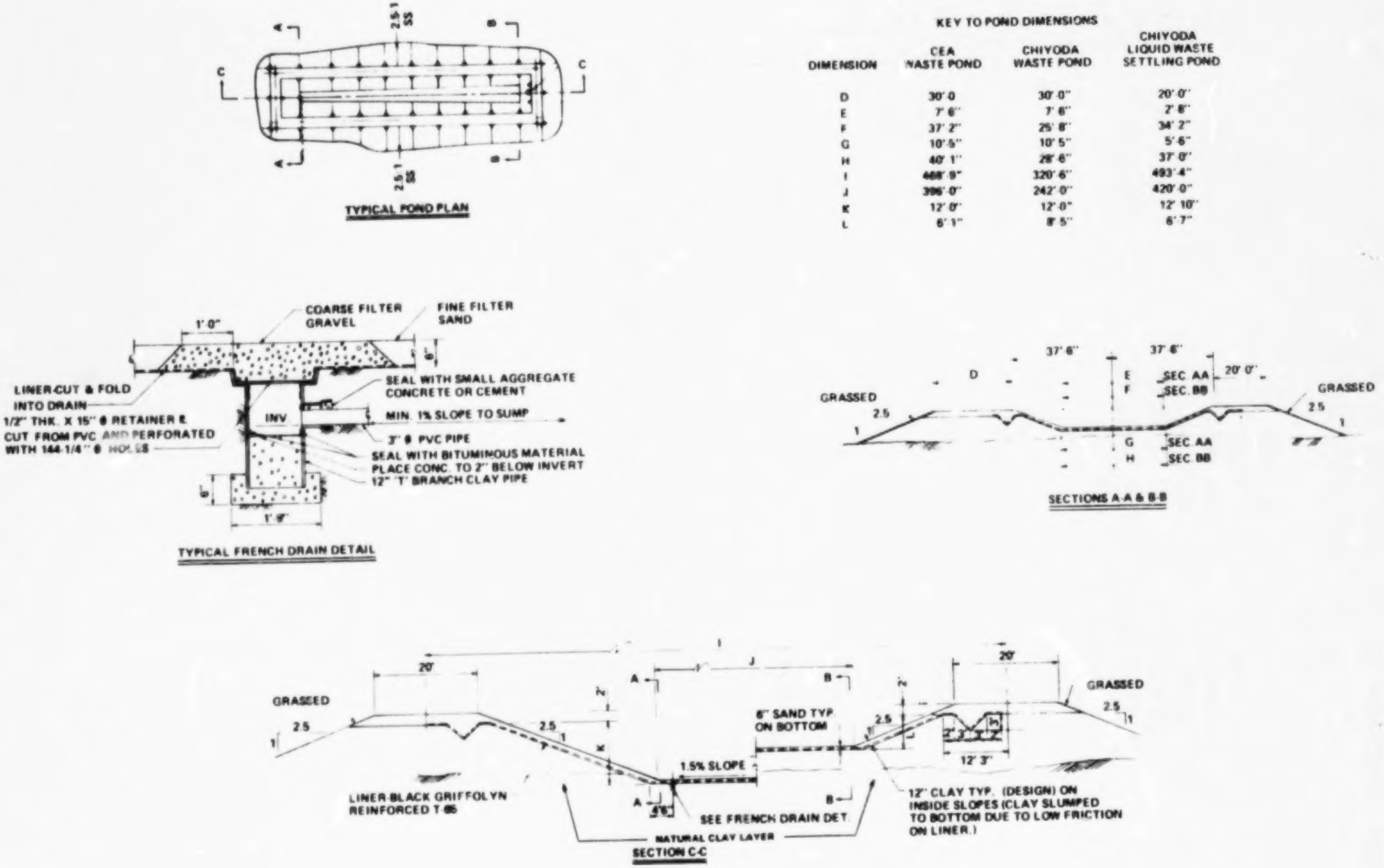


FIG. 1.3-9. CEA/ADL Dual Alkali Process Waste Pond


A
B
C
2.5:1
SS
TYPICAL POND PLAN
COARSE FILTER GRAVEL
FINE FILTER SAND
1' 0"
6"
LINER-CUT & FOLD INTO DRAIN
1/2" THK. X 15" Ø RETAINER E CUT FROM PVC AND PERFORATED WITH 144-1/4" Ø HOLES
SEAL WITH SMALL AGGREGATE CONCRETE OR CEMENT
INV
MIN. 1% SLOPE TO SUMP
3" Ø PVC PIPE
SEAL WITH BITUMINOUS MATERIAL
PLACE CONC. TO 2" BELOW INVERT
12" 'T' BRANCH CLAY PIPE
1' 9"
TYPICAL FRENCH DRAIN DETAIL
GRASSED
2.5
1
D
37' 6"
37' 6"
E
F
SEC. AA
SEC. BB
20' 0"
G
H
SEC. AA
SEC. BB
SECTIONS A-A & B-B
I
J
20'
20'
GRASED
GRASSED
2.5
1
L
K
A
B
6" SAND TYP ON BOTTOM
1.5% SLOPE
LINER-BLACK GRIFFOLYN REINFORCED T-65
4'6"
SEE FRENCH DRAIN DET.
NATURAL CLAY LAYER
12" CLAY TYP. (DESIGN) ON INSIDE SLOPES (CLAY SLUMPED TO BOTTOM DUE TO LOW FRICTION ON LINER.)
12' 3"
SECTION C-C

KEY TO POND DIMENSIONS

DIMENSION	CEA WASTE POND	CHIYODA WASTE POND	CHIYODA LIQUID WASTE SETTLING POND
D	30'-0	30'-0"	20'-0"
E	7' 6"	7' 6"	2' 8"
F	37' 2"	25' 8"	34' 2"
G	10'-5"	10' 5"	5'-6"
H	40' 1"	28'-6"	37' 0"
I	468'-9"	320'-6"	493'-4"
J	396'-0"	242'-0"	420'-0"
K	12' 0"	12'-0"	12' 10"
L	6' 1"	8' 5"	6'-7"

FIG. 1.3-10 Details of CEA/ADL and CT-101 Waste Ponds and CT-101 Liquid Waste Settling Pond

TABLE 1.3-5

CEA/ADL DUAL ALKALI PROCESS-DRY ASH REQUIREMENTS FOR OBTAINING SOLIDS CONTENTS OF 70% AND 75% IN MIXTURES OF CEA/ADL WASTE[1] AND DRY FLYASH

CEA WASTE SOLIDS CONTENT AS PRODUCED	% ASH OF MIXTURE DRY WEIGHT REQUIRED TO OBTAIN 70% SOLIDS	% ASH OF MIXTURE DRY WEIGHT REQUIRED TO OBTAIN 75% SOLIDS
40%	71%	78%
45%	65%	73%
50%	57%	67%
55%	48%	59%
60%	36%	50%
65%	20%	38%

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[1] Also applicable to direct lime or limestone FGD waste.

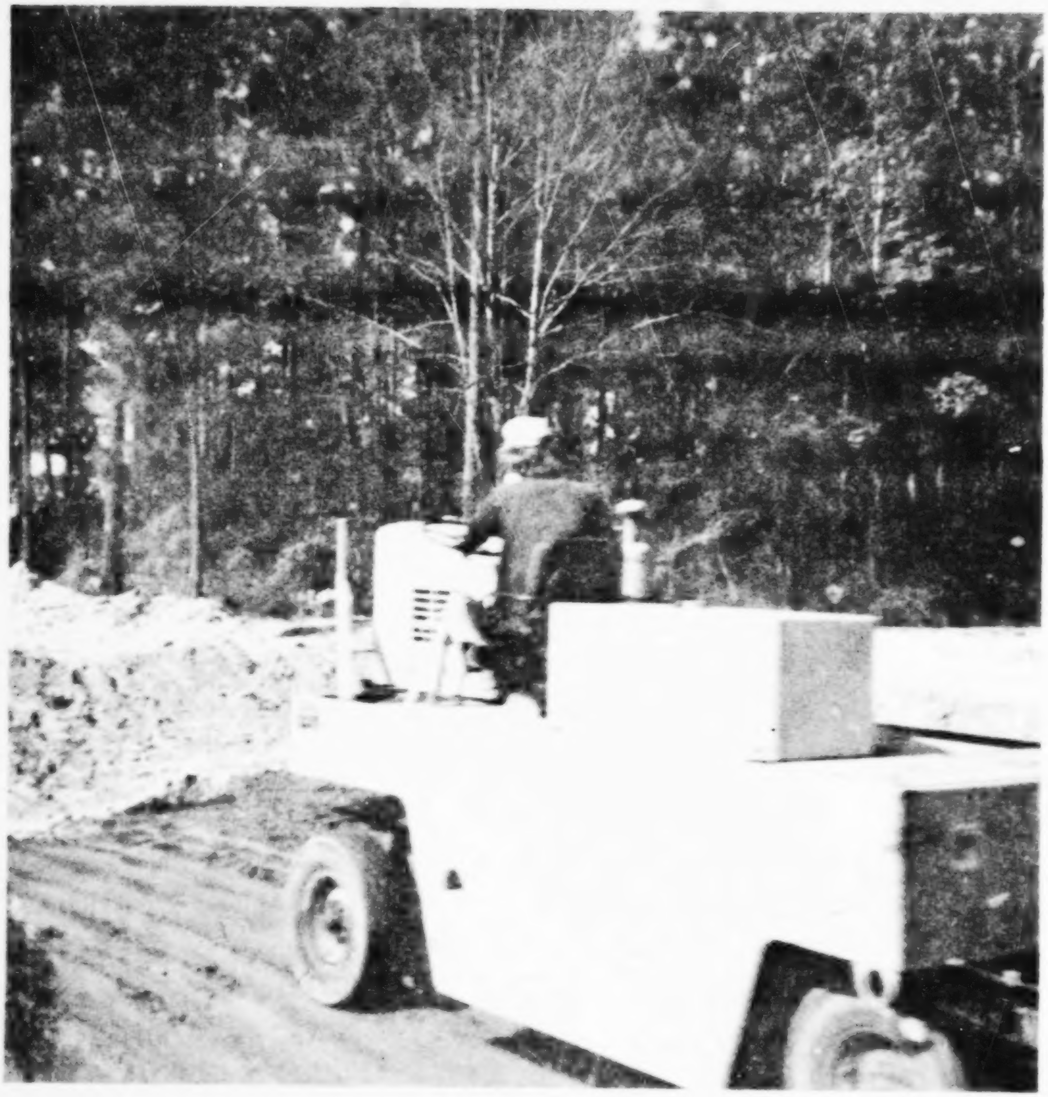

FIG. 1.3-11. CEA/ADL Dual Alkali Process-Compaction of CEA Waste, Dry Flyash, and Lime Mixture

TABLE 1.3-6

CEA/ADL DUAL ALKALI PROCESS VARIATION OF SOLIDS CONTENT AND DRY DENSITY IN LANDFILLED AND COMPACTED MIXTURES OF CEA/ADL WASTE AND DRY FLYASH

SAMPLE LOCATION	RANGE OF SOLIDS CONTENTS (%)	RANGE OF DRY DENSITIES (lb/ft^3)
First Lift Without Lime	75.5 - 77.2	82.5 - 87.8
Second Lift Without Lime	72.5 - 72.7	73.7 - 75.5
First Lift With Lime	72.5 - 76.5	76.9 - 80.2
Second Lift With Lime	77.2 - 80.1	73.8 - 81.0

indicated that adequate bearing strength and traction for operation of conventional earth moving equipment could be obtained in a landfill of CEA/ADL waste and dry flyash or CEA/ADL waste, dry flyash, and lime. No tendancy toward instability or reslurrying with rainfall was observed. Laboratory consolidation testing indicated that short-term settlement in the fill will be low, but that long-term settlement (creep) may pose constraints on future structural use of the landfilled area.

The effect of adding lime (about 1% dry weight basis) to a mixture of CEA/ADL waste and dry flyash was evaluated by laboratory and field testing. In field tests at Scholz, the addition of lime appeared to slightly increase the strength of the fill and to reduce the quantity of efflorescence (soluble salts coming to the surface of the fill as water evaporates). IU Conversion Systems (a vendor of FGD waste disposal technology) conducted laboratory tests on samples of CEA/ADL waste mixed with dry flyash and lime. They report that in addition to increasing strength and reducing efflorescence, the addition of lime prevents saturation of the landfill with water and therefore prevents permeation of leachate through the fill. Even if lime is not added to the landfilled mixture, the quantity of leachate seeping through the landfill will be less than that seeping through a pond due to the lower hydraulic gradient in the fill (assuming the permeabilities and waste thicknesses are equal).

(2) Ponding - If the CEA/ADL waste is repulped after vacuum filtration and slurried to a pond, laboratory testing corroborated by information reported in the literature[3] indicates that the settled waste in the pond would have a dry density of 30 to 45 pounds per cubic foot. (Vacuum filtration would be required even if the waste is slurried to a pond due to the need to wash sodium out of the waste solids prior to disposal.) Laboratory testing also indicated that the slurry of CEA/ADL waste solids would be deposited in a pond at a slope similar to that at which slurried flyash is deposited in a pond.

B. Utilization Potential - The potential for the utilization of CEA/ADL waste as a soil amendment for agricultural purposes was evaluated with greenhouse testing. CEA/ADL waste was mixed with soil at dosages ranging from less than 1 ton per acre to 20 tons per acre and leachate analysis and plant tolerance testing were conducted. The testing indicated that the waste provides calcium and sulfur in a form usable by the plants and can be applied at up to 10 tons per acre with no deleterious effects on plants or groundwater. However, significant problems are anticipated in spreading the waste on crop land unless it is dewatered beyond the solids content typical at Scholz* or unless major modifications are made to conventional agricultural spreading equipment. Nevertheless, it is doubtful that agricultural or any other utilization of a significant quantity of the CEA/ADL waste can be accomplished.

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* The range of CEA/ADL waste solids contents acceptable for conventional agricultural spreading has not yet been quantified.

1.3.4.6.4 Leachate and Runoff

A. Monitoring - The pond in which the CEA/ADL waste at Scholz was disposed was equipped with a sand drainage blanket sloping to an underdrain. The drainage blanket and underdrain details are shown in Figure 1.3-10. Effluent from the underdrain, which had a flow rate of less than 1 gpm, was monitored monthly for trace elements and major constituents, Table 1.3-7. The pH of all monthly underdrain samples and the suspended solids concentration in two monthly samples were outside the acceptable range of EPA effluent limitations for the area runoff category* of Steam Electric Power Generating Point Sources[4]. One or more of the monthly samples were found to be outside the acceptable range specified by EPA criteria for water supply[5,6] for pH, arsenic, boron, cadmium, chromium, iron, manganese, mercury, selenium, chloride, and sulfate. It is important to note that there were no violations of water quality standards or criteria in the receiving waters (where they are applicable) due to the CEA/ADL underdrain effluent at Scholz.

B. Concentration Reduction with Time - The sulfate, chloride, and total dissolved solids in leachate or runoff from a CEA/ADL waste disposal area will decrease with time as indicated by the results of five consecutive 48-hour shake tests shown in Table 1.3-8. The concentrations of sulfate, chloride and total dissolved solids in the wash water decreased by about a factor of 10 in going from the first shake test wash to the fifth shake test wash. Although the shake tests were conducted according to a standardized procedure followed by IU Conversion Systems, only a qualitative conclusion can be drawn (i.e. the leachate or runoff concentrations will decrease with time).

C. Source of Contaminants - Concentrations of arsenic, chloride, fluoride (which was found in excess of U.S. Public Health Service Drinking Water Standards[7] in the pond underdrain), mercury, and selenium were determined for all inputs to the CEA/ADL process. The analysis indicated that greater than 95% of these constituents entering the process originated in the coal.

D. Minimizing Surface and Groundwater Contamination - The only proven means of minimizing contamination of surface or groundwater from FGD waste leachate and runoff is to minimize the flow of liquid leaving the FGD waste. The leachate flow can be minimized by (1) preventing saturation of the waste and tying up the liquid initially present in the waste; (2) reducing the permeability of the waste and maintaining a low hydraulic gradient above the waste; or (3) placing a very low permeability material (or taking advantage of a natural layer) either above or below the waste. Options (1) and (2) would be applicable to landfill only, while option (3) would be applicable to either landfill or ponding. In addition, surface runoff from a landfill or supernatant overflow from a pond can be minimized through proper drainage design if the overall site and process water balance is maintained. The CEA/ADL process at Scholz was always operated closed loop with respect to the process itself. However, an evaluation of the water balance of any nonregenerable FGD system must also include the waste disposal area. As previously mentioned, this introduces several site specific considerations into the analysis.

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* The area runoff limitations by court order have been set aside and remanded to EPA for study and possible revision.

TABLE 1.3-7

SCHOLZ FGD PROJECT
EFFLUENT MONITORING PARAMETERS

Temperature

pH

Conductivity

Dissolved Oxygen

Total Dissolved Solids

Total Suspended Solids

Total Calcium

Total Magnesium

Total Sodium

Total Potassium

Total Hardness

Total Phosphorous

Dissolved Silica

Sulfate

Sulfite

Carbonate

Bicarbonate

Hydroxide

Chloride

Carbon Dioxide

Total Acidity

Color

Turbidity

Total Fluoride

Total Aluminum

Total Arsenic

Boron

Total Cadmium

Total Chromium

Total Copper

Total Iron

Total Lead

Total Maganese

Total Mercury

Total Nickel

Total Selenium

Total Zinc

Oil & Grease

Nitrate

Chemical Oxygen Demand

TABLE 1.3-8

CEA/ADL DUAL ALKALI PROCESS-
ANALYSIS FROM FIVE CONSECUTIVE 48-HOUR SHAKE TESTS CONDUCTED ON THE CEA/ADL WASTE SAMPLE AS PRODUCED

PARAMETER	WASH WATER CONCENTRATION (mg/l EXCEPT pH)				
	FIRST WASHING	SECOND WASHING	THIRD WASHING	FOURTH WASHING	FIFTH WASHING
pH	12.4	12.2	12.0	11.8	11.2
Phenolphthalein Alkalinity as $CaCO_3$	1700	970	600	460	340
MO Alkalinity (Total) as $CaCO_3$	1860	1080	700	530	370
Hardness as $CaCO_3$	810	380	550	530	520
Sulfite as SO_3	40	20	15	15	36
Sulfate as SO_4	2895	1168	471	250	181
Chloride as Cl	66	31	6	16	6
Total Dissolved Solids	6238	2736	1434	1014	682
Total Suspended Solids	2494	212	511	28	44

1.3.4.7 Operating Costs

1.3.4.7.1 Electrical Energy Consumption - For a CEA/ADL concentrated-mode, lime-regeneration dual alkali system incorporating a tray tower, and removing about 95% of the sulfur dioxide produced from burning 3 to 4 wt% sulfur coal, electrical energy consumption will be about 1.3% of station generation. For a system configured to remove particulate, and including both a tray tower and a variable throat venturi operating at about 12 inches of water pressure drop, the energy consumption will be approximately 2.5% of station generation. This assumes operation on a boiler producing about 2100 scfm of flue gas per megawatt of electrical generation and a ponding concept of waste disposal, including flyash and FGD waste, both of which are sluiced in a common pipe to the pond. Although other waste disposal configurations might be used (notably landfill waste disposal), this is one of the more typical physical arrangements that might be employed in a full-scale system.

1.3.4.7.2 Reheat - The total site energy consumption of any wet FGD process is the sum of electrical energy as well as the thermal energy for reheat (if any). Flue gas reheat in the CEA/ADL system (and the CT-101 system) at Scholz was by direct combustion of No. 2 fuel oil; however, the system was usually operated without reheat in order to conserve oil.

The questions of how much energy is required for reheat (or indeed if reheat is necessary) and which of the various possible methods is best are far too complicated to be addressed in a test program such as the one at Scholz. These choices are affected by a variety of legal as well as technical constraints such as avoidance of condensation and/or visible plume, dispersion of residual sulfur oxides, and plant location (climate and meteorology). The balance of the necessary trade-offs will have to be determined on a site-specific basis and will bear little, if any, relation to experiences with reheat systems at Scholz. If reheat is necessary, reheating flue gases 50°F will require from 1.4 to 2.1% of the energy input to a power station[8].

1.3.4.7.3 Lime Utilization - Utilization of the available calcium in the lime exceeded 95% as long as the residence times in the reactor system were similar to those that are now considered necessary by CEA/ADL for good utilization (3-5 minutes in the first reactor and 30-40 minutes in the second reactor). During operation on 3 to 4 wt% sulfur coal in April and May of 1976, utilization of available calcium was on the order of 93%. However, due to the higher feed forward rates required by the high inlet sulfur loading, reactor residence time was approximately two-thirds that anticipated for future designs. Operation with high particulate loading in the flue gas or highly fluctuating gas flows had no effect on utilization.

1.3.4.7.4 Sodium Makeup - In operations where the cake is thoroughly washed, sodium losses (as carbonate) will generally represent less than 5 mole % (8.3 wt%) of the sulfur dioxide absorbed and for a properly operated, closely monitored system, sodium losses on the order of 2 to 3 mole % can be achieved. This is, of course, in excess of any sodium lost in the pump seals, vessel overflows or as entrainment. However, a properly designed system would include a leak and spill collection system and sufficient liquor surge capacity in the process vessels to maintain leak and spill losses at less than 0.1 mole % (0.17 wt%) of the sulfur dioxide removal as sodium carbonate. In addition, entrainment losses measured at Scholz indicated that, for a properly operating mist eliminator, sodium losses are less than 0.1 mole % (as sodium carbonate) of the sulfur dioxide removed in the flue gas.

1.3.4.7.5 Other Operating Costs - It is not possible to reliably scale up the operational and maintenance requirements from such a small system to be quantitatively representative of the requirements of a full-scale system. However, the experience at Scholz indicates that overall maintenance requirements on a properly designed and operated concentrated-mode, lime-regeneration sodium/calcium dual alkali system will probably run somewhere between equal to and 25% less than similar costs on direct lime or limestone systems. Operational manpower requirements will be similar to that of direct lime or limestone systems.

1.4 CHIYODA THOROUGHBRED 101 (CT-101) Process

1.4.1 System Design - In the CT-101 process, Figure 1.4-1, flue gas passes through a prescrubber (at Scholz a fixed throat venturi, Figure 1.4-2) where flyash and chlorides are removed. From there it flows to a packed absorption tower, Figure 1.4-3, where sulfur dioxide and sulfur trioxide are removed. The sulfurous acid produced is oxidized to sulfuric acid in an iron-catalyzed reaction in an oxidation vessel, Figure 1.4-3. This sulfuric acid is partially neutralized with limestone in a specially designed crystallizer, Figure 1.4-4, to produce gypsum. The gypsum is separated from the partially neutralized acid -- mother liquor -- (by centrifuge at Scholz, Figure 1.4-5), and trucked to a disposal pond (at Scholz), while the mother liquor is recirculated to the absorption section.

The Chiyoda process operates at such a low liquid pH (less than 1) that the potential of failure in liners that protect carbon steel vessels from corrosion is a problem. Because of this, the process vessels in the absorption/crystallization section at Scholz were constructed from 316L stainless steel or from fiberglass. However, chloride ions in excess of about 400 ppm produce pitting corrosion in 316L in the CT-101 process's chemical system. Because of this, the venturi prescrubber was installed ahead of the absorption/crystallization sections of the process to minimize chloride (as well as particulate) collection in these later portions of the system.

Under the best of circumstances, some chlorides will pass through the prescrubber and be captured in the absorber, causing the chloride concentration in the absorption/crystallization sections to build to a level sufficient to maintain a material balance. If this natural level is above about 400 ppm, it is necessary to institute a liquid purge from the system for chloride control. However, at Scholz, the process requirements for water input into the absorption/crystallization section were in excess of the water which left the system associated with gypsum cake*. (This was due principally to the excessive flushing water requirements of the mechanical seals used on the oxidizer feed pumps.) Therefore, a free liquid blowdown was necessary to maintain the water balance in the absorption/crystallization portion of the process exclusive of any chloride purge requirements.

To accomplish this, a bleed stream was taken from the mother liquor return line. During early operations, this bleed stream was combined with a bleed from the prescrubbing section and pumped to a neutralization tank. Limestone was added to this neutralization tank to bring the pH up close to neutrality (6 to 8), and the combined slurry of ferric hydroxide (from the catalyst), unreacted limestone, gypsum, and neutralized liquor overflowed into a settling pond. The clear supernatant from this settling pond was fed into the plant ash pond.

---

* Initially there were no evaporative losses from the absorption/crystallization section since flue gas saturation took place in the prescrubber which was isolated on the liquid side from the remainder of the process. The water-savings modifications discussed later joined these sections of the process together on the liquid side.

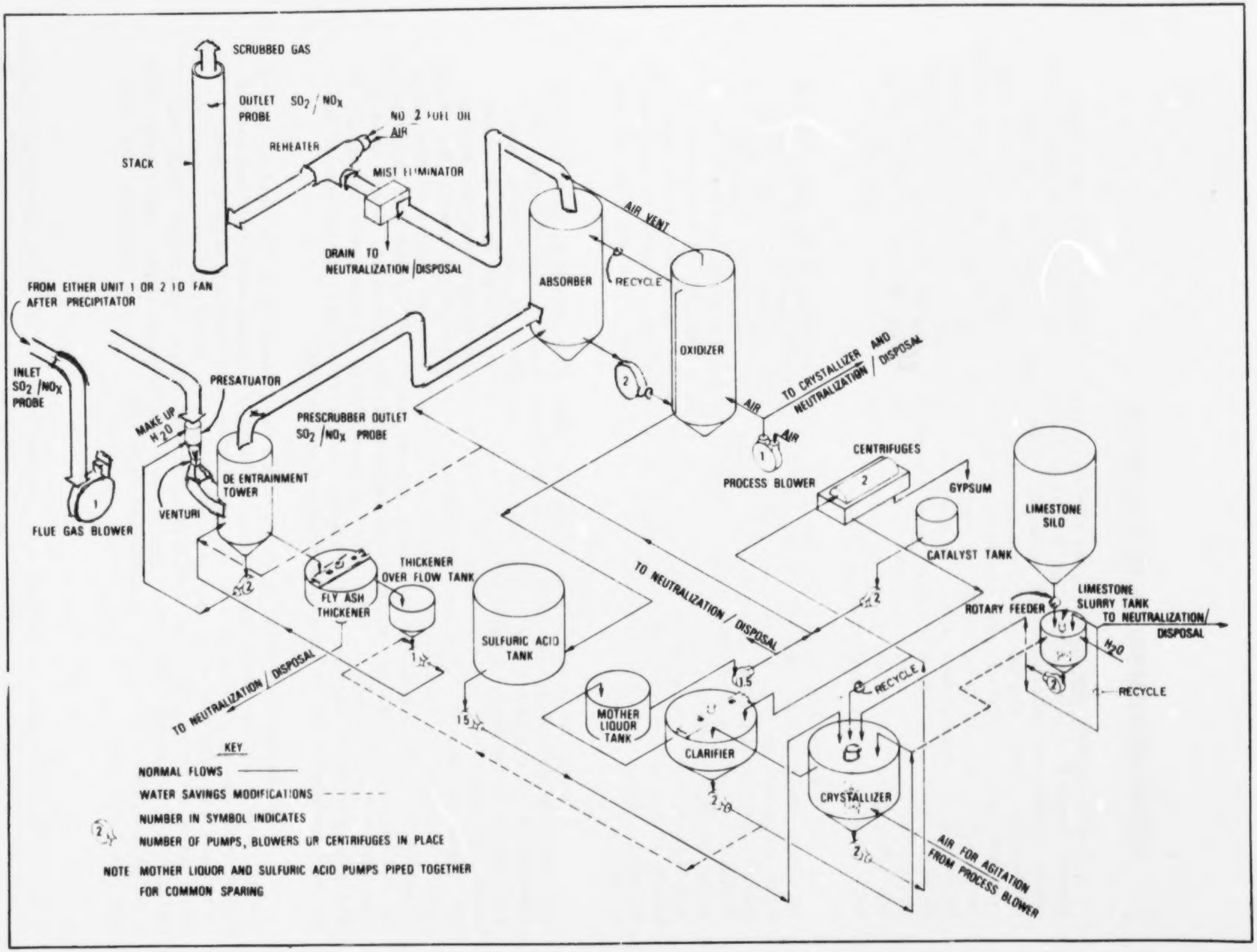
SCRUBBED GAS
OUTLET SO_2/NO_x PROBE
STACK
REHEATER
NO 2 FUEL OIL
AIR
MIST ELIMINATOR
DRAIN TO NEUTRALIZATION / DISPOSAL
AIR VENT
ABSORBER
RECYCLE
OXIDIZER
FROM EITHER UNIT 1 OR 2 I.D. FAN AFTER PRECIPITATOR
INLET SO_2/NO_x PROBE
PRESATUATOR
MAKE UP H_2O
PRESCRUBBER OUTLET SO_2/NO_x PROBE
DE ENTRAINMENT TOWER
VENTURI
FLUE GAS BLOWER
1
2
AIR
TO CRYSTALLIZER AND NEUTRALIZATION / DISPOSAL
AIR
PROCESS BLOWER
CENTRIFUGES
GYPSUM
CATALYST TANK
LIMESTONE SILO
LIMESTONE SLURRY TANK
ROTARY FEEDER
TO NEUTRALIZATION / DISPOSAL
H_2O
RECYCLE
THICKENER OVER FLOW TANK
FLY ASH THICKENER
SULFURIC ACID TANK
TO NEUTRALIZATION / DISPOSAL
TO NEUTRALIZATION / DISPOSAL
1.5
MOTHER LIQUOR TANK
CLARIFIER
CRYSTALLIZER
RECYCLE
AIR FOR AGITATION FROM PROCESS BLOWER
KEY
NORMAL FLOWS
WATER SAVINGS MODIFICATIONS
NUMBER IN SYMBOL INDICATES
NUMBER OF PUMPS, BLOWERS OR CENTRIFUGES IN PLACE
NOTE MOTHER LIQUOR AND SULFURIC ACID PUMPS PIPED TOGETHER FOR COMMON SPARING

FIG. 1.4-1. CT-101 Process Flow Diagram

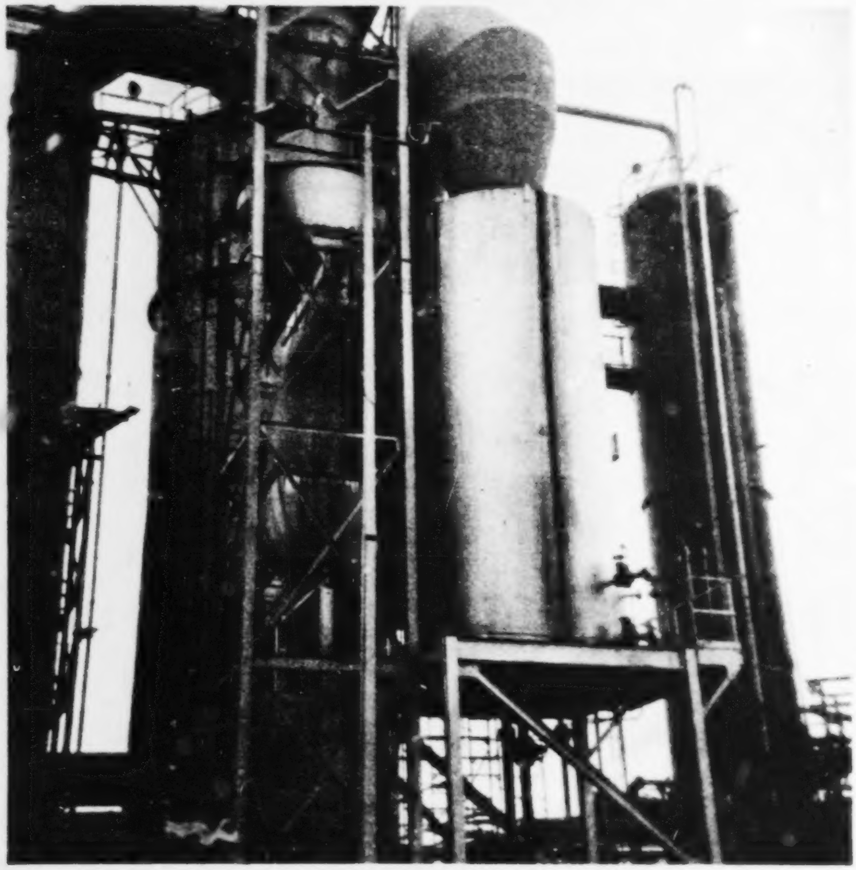

FIG. 1.4-2. CT-101 Process
Venturi and Entrainment Separation Tower

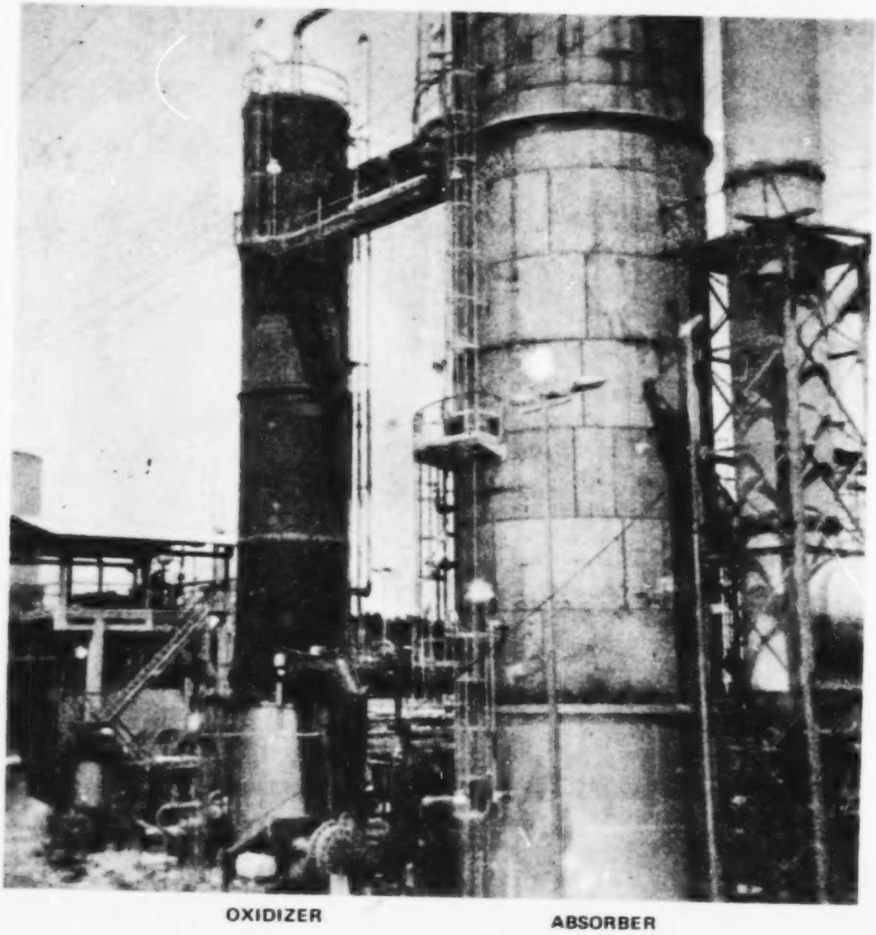

FIG. 1.4-3. CT-101 Process Absorber and Oxidizer

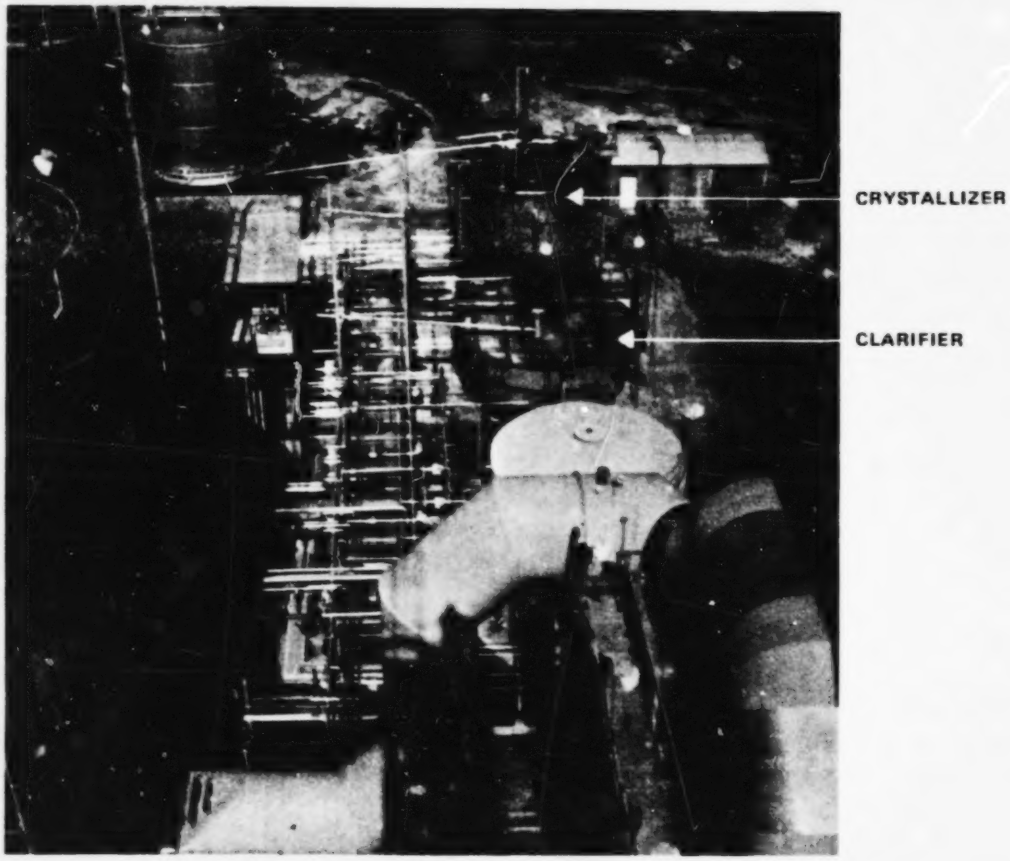

FIG. 1.4-4. CT-101 Process
Crystallizer and Clarifier

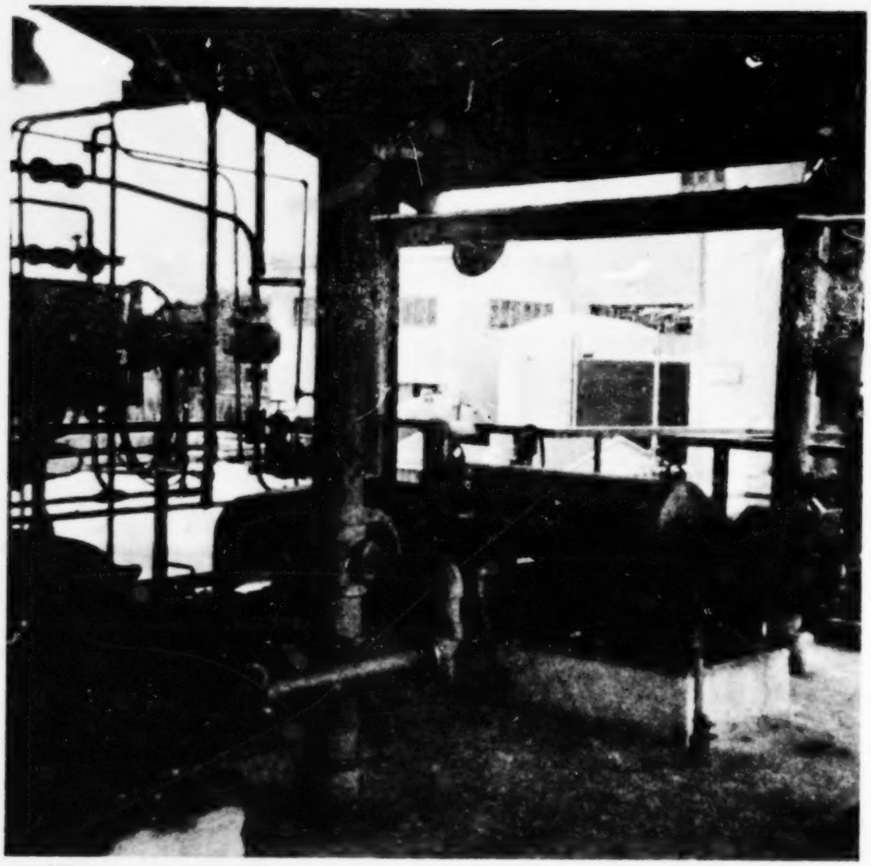

FIG. 1.4-5. CT-101 Process Centrifuge

The neutralization system as designed at Scholz is not typical of what would be used on a completely integrated, full-scale system. Because of the small size of the process at Scholz, Chiyoda made the decision not to install a waste-neutralization, catalyst-recovery system as they have on large systems in Japan. According to Chiyoda, this recovery system essentially involves the addition of a clarifier at the neutralization tank to recover precipitated catalyst, unreacted limestone, and gypsum from the bleed stream prior to its discharge.

Since the process, as it was originally designed, could not be operated in a mode which limited the blowdown of free liquid to levels which were thought technically feasible, it was modified during 1976. This revision work is referred to as "water-savings modifications" and the testing of these modifications is referred to as "Water-savings testing." These modifications (shown as dashed lines in Figure 1.4-1) allowed (1) the addition of acidic clarifier underflow liquor (containing gypsum seed crystals) to the limestone feed preparation tank instead of fresh water, (2) the addition of mother liquor bleed as makeup to the prescrubber section and (3) the addition of clarifier underflow to the prescrubber section (should seed crystals be necessary for scale control there.)

1.4.2 Process Chemistry - In the absorption section of the CT-101 process, absorption and oxidation of sulfur dioxide occur to produce a dilute solution of sulfuric acid according to the overall reaction:

$$SO_2 + \tfrac{1}{2}O_2 + H_2O \xrightarrow{Fe^{+++}} H_2SO_4 \tag{1}$$

In addition, any sulfur trioxide which is absorbed is converted to sulfuric acid via the following reaction:

$$SO_3 + H_2O \longrightarrow H_2SO_4 \tag{2}$$

Although the first reaction -- the reaction of dissolved sulfur dioxide with oxygen in an aqueous solution -- has been reported by many authors, and various mechanisms have been reported[9,10,11], it is not fully understood. One mechanism that has been proposed, which involves a change in the oxidation state of the catalyst, ferric sulfate, from ferrous sulfate to ferric sulfate and back to ferrous sulfate again, is described as follows:

$$2FeSO_4 + SO_2 + O_2 \longrightarrow Fe_2\ (SO_4)_3 \tag{3}$$

$$Fe_2\ (SO_4)_3 + SO_2 + 2H_2O \longrightarrow 2FeSO_4 + 2H_2SO_4 \tag{4}$$

The absorption of sulfur dioxide in a chemical system such as this is inversely proportional to the acid concentration. Therefore, it is necessary to neutralize the acid at a rate equivalent to that at which it is being produced in order to maintain high sulfur dioxide removal efficiency. To accomplish this, a slipstream of the dilute sulfuric acid (less than 3 wt%) is taken to a crystallizer where limestone is added, the acid is neutralized, and gypsum is produced according to the following overall reaction:

$$H_2SO_4 + CaCO_3 + H_2O \longrightarrow CaSO_4 \cdot 2H_2O + CO_2\uparrow \tag{5}$$

The key to the successful operation of the system is the control of this reaction in the crystallizer by preferential precipitation on gypsum seed crystals rather than on the crystallizer internals.

1.4.3 System Operation - The system became operational on February 11, 1975, and was operated over a period of 25 months through March 22, 1977. During this period, it logged over 10,600 hours of operation with shutdown periods of varying length for maintenance, modifications, repairs, and when the boiler was shut down. The 25 months of operation can be broken down into five discrete periods, Table 1.4-1, defined by flue gas composition, coal characteristics, and system operation.

During the first operating period, covering the first six months of operation, the system operation was erratic mainly because of several unexpected mechanical problems. Sulfur dioxide concentrations in the inlet flue gas averaged approximately 1050 ppmv as compared to a design concentration of 2200 ppmv. Removal efficiency was generally in excess of 85%. The actual sulfur dioxide concentrations varied over a range of 600 to 1550 ppmv, often fluctuating daily and even hourly over the entire range of concentrations. However, unlike direct lime or limestone systems and dual alkali processes, the chemistry of the CT-101 process is not sensitive to inlet sulfur dioxide concentrations, and these low inlet values had no effect on process operations.

Prior to the second operating period, slightly higher sulfur coal was delivered in August, 1975. Use of this coal resulted in inlet sulfur dioxide concentrations that ranged between 800 and 1700 ppmv and averaged about 1250 ppmv. Sulfur dioxide removal averaged 83%. This period of operation was terminated on January 20, 1976, when a fiberglass reinforced plastic (FRP) discharge line on one of the oxidizer feed pumps, which had been damaged a few days earlier by a water hammer, broke and fell to the ground. Repairs and modifications required approximately five weeks.

The third operating period began when the system was restarted on February 26, 1976, and continued for six weeks through April 4, 1976. During three weeks of this period, operation was on high-sulfur coal, containing about 3.5 wt% sulfur, which provided inlet sulfur dioxide concentrations ranging from 1900 to 2600 ppmv and averaging about 2200 ppmv. Sulfur dioxide removal remained acceptable, generally exceeding 90%. This period of operation was terminated when the system was shut down so the water-savings modifications could be made. About a week before the process was ready to restart, the FRP oxidizer vessel caught fire from welding slag. Repairs required slightly over three months and it was not until July 28, 1976, that the process was restarted.

The fourth operating period, which spanned three months, commenced on July 28, 1976. During this period, the water-savings modifications were tested along with the impact of high inlet particulate loadings and fluctuating gas loads. Sulfur dioxide concentrations in the inlet flue gas ranged between 1000 and 2200 ppmv and averaged about 1300 ppmv. Sulfur dioxide removal generally exceeded 80%. The fourth operating period ended with the termination of formal testing on November 1, 1976.

TABLE 1.4-1

CT-101 PROCESS
SUMMARY OF OPERATING PERIODS

PERIOD	INTERVAL	OPERATING HOURS	COAL SULFUR	TESTING
1	2/11/75-8/3/75	1980	Low	Start-up and Shakedown
2	9/15/75-1/20/76	2940	Low/Medium	Stable Load
3	2/26/76-4/4/76	893	High	Stable Load and High Sulfur
4.	7/28/76-11/1/76	1444	High and Low/Medium	Fluctuation Load, Water Savings, and Particulate
5[1]	11/2/76-3/26/77	3366	Low/Medium	Reliability

---

[1] Operating period 5 is not analyzed in detail in this report since the formal Southern Company test program ended after period 4.

After the end of the formal test program, the system continued to operate for a period of almost five months (period 5) until March 22, 1977. At that time operation was terminated to allow Chiyoda to modify the CT-101 process to a new forced-oxidation limestone process they have developed (CT-121)[12]. Operation of this new system, which will be by Chiyoda, is expected to begin by mid-1978. Negotiations are currently underway to allow EPRI to conduct a detailed evaluation of the new system.

1.4.4 System Performance

1.4.4.1 Reliability/Operability Potential

The CT-101 process has been operating in several large-scale installations in Japan for several years; and, on the whole, its reliability record there has been impressive. However, all of the processes installed on utility boilers in Japan are in service on combustion products produced by firing oil. Because of the small system size (20 megawatts nominal), and the fact that this was the first application of the CT-101 technology on combustion products from coal-fired boilers, the system at Scholz has been considered a prototype for the purposes of this evaluation.

In an effort to put the operation of the process on a clearly visible basis, the 4 PEDCo FGD system viability parameters[2] have been calculated for the system. These parameters are represented as cumulative percentages in Table 1.4-2. In addition, the operability parameter on a monthly basis is presented in Figure 1.4-6.

As was previously mentioned, none of these parameters is adequate for evaluating the future performance potential of prototype systems. Even for full-scale systems, they cannot be used literally to judge an entire technology by results at only one site. They are, however, the only quantitative measures of FGD system operation that approach universal acceptance and for that reason have been presented here. What is needed for a proper evaluation is long-term operation on several large, base-loaded boilers. Lacking that, a qualitative judgment of the system's potential for acceptable long-term operation is the only means of evaluation available.

1.4.4.1.1 Process Performance - Process operability (ease of operation) was excellent in all respects. The system was successfully operated over a range of widely fluctuating inlet flue gas sulfur dioxide levels, oxygen concentrations, and flow rates with little change in the sulfur dioxide removal efficiency, waste solids properties, or limestone utilization. It demonstrated an ability for continuous operation with large and frequent variations in absorbent and mother liquor acid concentrations and limestone and gypsum slurry concentrations.

The most impressive aspect of the system's operation is its resistance to short-term upsets. The large liquid holdup acts much like a capacitor and absorbs most of the short-term process variations. In addition, because of the large volume of gypsum seed crystals in the crystallizer, the likelihood of calcium/sulfur salt deposition in the system is so small as to be almost negligible. Furthermore, operation of the system for the entire test program (over

TABLE 1.4-2

CT-101 PROCESS[1]
CUMULATIVE VIABILITY PARAMETERS
(FEBRUARY 1975-MARCH 1975)

PARAMETER	VALUE (%)	TOTAL TIME NUMERATOR (HOURS)	TOTAL TIME DENOMINATOR (HOURS)
Availability	81	12,870	15,920
Reliability	75	10,623	14,180
Operability	70	10,623	15,206
Utilization Factor	68	10,623	15,920

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[1] The period of fire damage repairs is excluded from thie analysis.

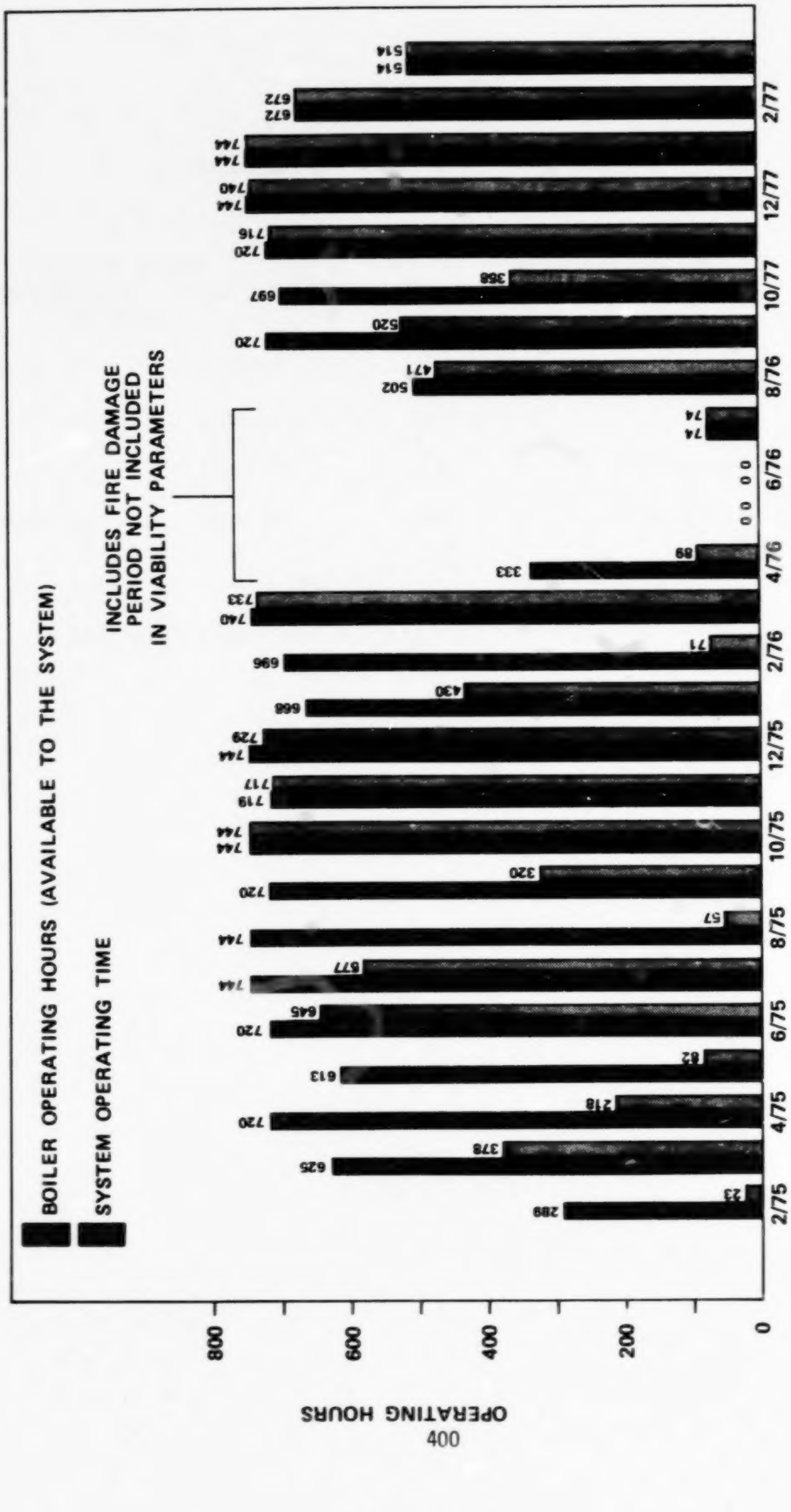

FIG. 1.4-6. Operability of the CT-101 Process

10,600 hours), without mist eliminator washing and without any mist eliminator scaling, confirmed the viability of system operation without a mist eliminator wash.

As previously mentioned, the process, as originally designed, had a limitation in that it could not be operated with a "closed-loop" water balance in the same manner as was done with the CEA/ADL system. However, after the water-savings modifications it was operated for a period of one month with the total liquid discharge (the sum of the blowdown plus the moisture in the gypsum cake), nominally the same as that which would leave in a 50% solids waste cake from a calcium-sulfite-producing FGD system*. However, there was still a free liquid discharge from the prescrubber blowdown. During the water-savings testing this stream was reduced to 3-5 gpm (0.15-0.25 gpm/MW) and can probably be further reduced. The treatment or disposal of this stream will be a site-specific consideration. For example, if the gypsum produced at the site were mixed with dry flyash and deposited in a landfill, this stream could be disposed of in the ash/gypsum mixture. Under other circumstances, it might be disposed of in a pond along with the gypsum or some type of treatment might be necessary. The total volume is low enough that several alternatives are feasible.

As has previously been mentioned, it is difficult to generalize with respect to the long-term reliability potential of any mechanical/chemical system. However, the operation of the CT-101 system at Scholz and elsewhere indicates that the overall performance of a properly designed and operated CT-101 system should be superior to that of direct lime and limestone systems because;

- The system is highly resistant to upset. The potential for scaling is eliminated, except in extreme upset conditions, by the large volume of gypsum seed crystals present in the crystallizer. Calcium/sulfur salt precipitation occurs on these crystals preferentially because of their large surface area.
- The handling of slurries in the absorption section is completely eliminated.
- Unlike lime and limestone systems, which must control pH closely, pH is not a control parameter in the CT-101 system except in a secondary sense to acid concentration. Close control of acid concentration is not necessary to control scaling.

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* The use of the term "a 50% solids waste cake from a calcium-sulfite-producing FGD system" is somewhat arbitrary. It represents the lower-bound of ash-free filter cake solids content that is typical of the CEA/ADL dual alkali process waste and the upper-bound for direct lime or limestone waste (assuming the latter is filtered). The intent of the comparison is to show that the total liquid discharged from the CT-101 process at Scholz, after modification to the liquor loop, was essentially the same (and in some cases lower) than that discharged from similar wet FGD systems.

1.4.4.1.2 Equipment Performance - As with the dual alkali system, the overall mechanical performance of the CT-101 system was generally acceptable. All but a few of the problems that developed during the course of the program, as discussed below, were resolved by simple operational adjustments or equipment modifications.

A. Mechanical Problems - Aside from occasional failures of rubber-lined pumps, the most significant equipment problems in the system involved the centrifuges, the gas-side expansion joints, the synthetic liner in the venturi throat, and failure of fiberglass in the oxidizer and associated piping. In addition, there was a serious fire in the oxidizer (fiberglass construction). These problems accounted for the majority of the mechanical-related downtime and maintenance.

(1) Aside from the problems associated with the fiberglass oxidizer vessel, the centrifuge (continuous screw-decanter, solid-bowl type) was the greatest source of trouble in the system. The major problems were high vibration and excessive scroll erosion. After replacement of a stainless-to-stainless running surface with a bronze-to-stainless surface, repairs to a cracked tub flange, and slowing of the scroll from 1400 to 800 G's (consistent with Chiyoda's experience in Japan), the vibration problem was generally resolved and scroll life stabilized at about 2000 hours.

Chiyoda's experience in Japan has generally shown basket-type centrifuges to be superior to the solid-bowl machines used at Scholz. (Solid-bowl centrifuges were used at Scholz because of size limitations.) In addition, tests conducted with a Bird-Young continuous-vacuum filter and a De Laval pusher centrifuge, near the end of the program, showed both to be superior to the solid-bowl centrifuge for this service.

(2) Most of the original gas side expansion joints (in both hot and saturated gas service) were manufactured from cotton duck. As these failed, they were replaced with Garlock viton and asbestos joints which performed well.

(3) The venturi was initially lined with an epoxy material (Caroline 505AR). While this liner appeared generally acceptable for the process conditions, it failed, from excessive temperatures, on an extension of the gas inlet duct located in the gas quench zone in the venturi. It was replaced with a high-temperature vinyl ester resin which also failed. (This latter material failed in a similar location in the CEA/ADL venturi as well.)

(4) Several problems, which resulted in over 1200 outage hours, occurred with field-erected, hand-laid fiberglass joints in the oxidizer and associated piping. After the joints were stripped out and strengthened, no further problems occurred.

(5) A fire started in the fiberglass oxidizer vessel when welding slag from a sample nozzle installation ignited the polypropylene gas/liquid distribution trays in the vessel. Damage repairs required over 2500 hours.

B. Instrumentation Problems - There were no instrumentation problems that were significant to process operation. However, the gypsum slurry density analyzer, a sonic type with in-line probes, failed after only a few weeks operation and was not replaced. Frequent manual samples were taken as an alternate.

1.4.4.2 Sulfur Dioxide Removal Capability

1.4.4.2 Sulfur Dioxide Removal Capability - The removal of sulfur dioxide in the CT-101 process is strongly a function of inlet sulfur dioxide concentration, absorbent liquid-to-gas ratio (L/G), absorbent sulfuric acid concentration, and absorbent catalyst concentration. Because of the high degree of interaction between these variables and the large liquid circulation rates required for adequate sulfur dioxide removal (design L/G's on the order of 400 gal/1000 acf are necessary for 90% removal of the sulfur dioxide produced from the combustion of 3 wt% sulfur coal), the ability to achieve a precisely desired sulfur dioxide removal efficiency is more limited in the CT-101 process than in a system like dual alkali.

The high liquid circulation rates make throttling control of the absorbent impactical. Because of this, the process at Scholz was configured so that the absorbent circulation rate was controlled by the number of oxidizer feed pumps, either one or two, in service at any time. Since the process energy consumption was significantly higher (see Section 1.4.4.7.1) with two pumps operating and high sulfur dioxide removal was not required, the process was normally operated with only one oxidizer feed pump in service. However, sulfur dioxide removals in excess of 90 to 95% could easily be achieved with the expenditure of more energy.

Average sulfur dioxide removal efficiencies for various phases of each operating period are tabulated in Table 1.4-3. Correlations of sulfur dioxide removal with important process parameters are not reported because of Chiyoda's stipulation that this information remain confidential. However, the only effect of increasing or decreasing fuel sulfur content is to require more or less energy to produce a desired level of removal.

1.4.4.3 Nitrogen Oxides Removal Capability - Very little work was done on the nitrogen oxides (NO_x) removal capability of the system. However, the small amount of data that was collected agreed with Chiyoda's experience in Japan that nitrogen oxides removal efficiency is on the order of 5 to perhaps, at most, 10%.

1.4.4.4 Particulate Removal Capability and the Impact of Particulates on System Operation

A. Low Inlet Loading Conditions - Results from tests conducted during January and March of 1976, when the precipitator was fully energized, are summarized in Table 1.4-4. The purpose of these first two test sets was to establish the approximate value of the inlet particulate loadings as a reference point for gas conditions during routine operations. These results indicate that the inlet particulate loadings, during periods when the precipitator was fully energized, was less than 0.1 gr/dscf.

TABLE 1.4-3

CT-101 PROCESS
SUMMARY OF SO_2 REMOVAL PERFORMANCE

PERIOD	INTERVAL	APPROXIMATE OPERATING HOURS	NO. OF OXIDIZER PUMPS OPERATING	SCRUBBER CONDITIONS L/G (gals/1000 Acf_{sat})			INLET SO_2 (ppmv)		AVG. REMOVAL EFFICIENCY (%)
				L/G (gals/1000 Acf_{sat})	H_2SO_4 (wt %)	Fe^{+++} (ppm)	RANGE	AVG.	
1	6/17/75-7/27/75	900	1	220-400	1.8%	2200	600-2000	1100	90
2	9/15/75-1/20/76	2940	1	220-365	1.9%	2275	800-1700	1250	83
3	3/2/76-3/12/76	240	1	220-260	1.3%	2200	700-1700	1200	90
	3/16/76-3/23/76	200	2	310-525	1.3%	2300	1800-2200	2000	91
	3/24/76-4/4/76	290	1	170-490	1.3%	2000	1200-2800	2200	90
4	8/1/76-9/5/76[1]	590	1	150-240	1.6%	2000	1000-2100	1400	84
	9/10/76-9/26/76[2]	400	1	155-355	1.7%	1900	1200-2400	2000	89
	10/17/76-10/31/76[3]	360	1	150-295	1.4%	2000	800-1800	1100	82

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[1] Water savings tests - no mother liquor concentration

[2] Fluctuating load/water savings tests - 30% mother liquor concentration

[3] Particulate tests

TABLE 1.4-4

CT-101 PROCESS
PARTICULATE REMOVAL TEST RESULTS[1]
(JANUARY & MARCH 1976)

TEST NO.	PRECIPITATOR SECTIONS IN SERVICE	PARTICULATE LOADINGS INLET (gr/dscf)	PARTICULATE LOADINGS OUTLET (gr/dscf)	REMOVAL EFFICIENCY (%)	GAS FLOW (% OF DESIGN)	NO. OF OXIDIZER PUMPS IN SERVICE
JANUARY						
1	ALL	0.009	0.006	33	105	1
2	ALL	0.012	0.007	41.7	105	1
3	ALL	0.050	0.002	96	105	1
4	ALL	0.045	-----	---	105	1
MARCH						
1	ALL	0.072	0.0097	86.5	100	2
2	ALL	0.072	0.0054	92.5	100	2
3	ALL	0.080	0.0043	94.6	100	2

---

[1] Venturi pressure drop between 8 and 10 inches of water.

Depending on the inlet loading, removal efficiencies ranged between 33 and 96%, with a venturi pressure drop between 8 and 10 inches of water. Too few tests were conducted during January and March to allow confidence in the particulate removal efficiency data shown in Table 1.4-4. However, the data do support the idea that because of the large interfacial area for mass transfer in the packed tower absorber the system has good potential as a polishing particulate remover.

B. High Particulate Loading Conditions - As with the CEA/ADL process, it was felt that long-term operation with high particulate loadings in the inlet flue gas was not necessary, since several FGD systems of similar design are operating with full particulate loadings. Because of this, the electrostatic precipitator was left fully energized until the last two weeks of the test program. At that time, various sections, Figure 1.3-8, were deenergized to observe the short-term effects of high particulate loadings on system performance.

No significant impact was noted for particulate loadings of approximately 3 gr/dscf. Particulate removal efficiency exceeded 99%, Table 1.4-5, for gas flows at or near design conditions and a venturi pressure drop of nominally 7 to 10 inches of water. The high particulate loading tests were conducted with the system operating in the high water consumption, open-loop mode that was typical of operations prior to and after September, 1976. Operation in a water-savings mode similar to that of September, 1976, and with high ash loadings into the system, could require operational modifications. For example, the injection of gypsum seed crystals into the venturi loop to control calcium sulfate supersaturation might be necessary, particularly if the flyash has a high calcium content.

In interpreting these results, it should be remembered that the venturi is followed by a packed tower (about 30 feet of packing height). The packed tower contributes to the particulate removal and to some extent "masks" the effect of venturi pressure drop.

1.4.4.5 Impact of Contaminants on System Operation - As previously discussed, the chlorides present in the coal have a significant impact on the materials of construction in the system and can significantly affect the water balance. According to CIC, significant corrosion does not occur in the 316L stainless steel used extensively in the construction up to about 400 ppm chloride in the liquor because of the corrosion inhibiting effects of ferric ion (from the catalyst) and nitrate ion[13,14,15,16] (from NO_x in the flue gas). During operation at Scholz the system operated on the flue gas produced from the combustion of 2 to 3 wt% sulfur and 0.1 wt% chloride coal. However, the chloride levels in the absorption/crystallization section did not exceed 150 ppm.

Chiyoda has reported[17] the effects of several organic as well as inorganic compounds on sulfur dioxide removal in the absorption section. With the exception of iron (from the catalyst) and calcium and magnesium (from the limestone), none of the common metal ions were found in the process liquor at Scholz in concentrations in excess of 30 ppm. In all cases, this was significantly below the levels previously found by Chiyoda to have no effect on the process operation.

TABLE 1.4-5

CT-101 PROCESS
PARTICULATE REMOVAL TEST RESULTS
(OCTOBER 1976)

TEST NO.	PRECIPITATOR SECTIONS IN SERVICE	PARTICULATE LOADINGS INLET (gr/dscf)	PARTICULATE LOADINGS OUTLET (gr/dscf)	REMOVAL EFFICIENCY %	GAS FLOW (% OF DESIGN)	NO. OF OXIDIZER PUMPS IN SERVICE	VENTURI[2] Δ P ("H_2O)
1	C[1]	2.93	0.050	98.3	95	1	7.5
2	C	2.39	0.016	99.3	95	1	7.5
3	C	2.35	0.017	99.3	95	1	7.6
4	C	2.98	0.028	99.1	95	1	8.5
5	C	2.99	0.020	99.3	95	2	6.4
6	C	2.01	0.024	98.8	75	1	5.2
7	All	0.184	0.010	94.6	95	1	9.5
8	C	2.89	0.027	99.1	95	1	10.0

---

[1] Precipitator section "C" only in service.

[2] In interpreting these results, it should be remembered that the venturi is followed by a packed tower of about 30 feet of packing height. The packed tower contributes to the particulate removal and to some extent "masks" the effect of venturi pressure drop.

Magnesium levels in the process liquor were generally on the order of 700 to 800 ppm during normal process operation, (limestone containing less than 2% magnesium carbonate was used). However, in December of 1976 approximately 270 tons of high magnesium limestone was inadvertently delivered to the process, causing dissolved magnesium levels to rise to 15,000 ppm. The only effects were a slight reduction in sulfur dioxide removal* and a decrease in the particle size of the gypsum which caused the gypsum solids content to fall about 1 to 3 wt% from typical values of about 85 wt%.

1.4.4.6 Secondary Environmental Impact - A was stated in Section 1.3.4.6, the "Secondary Environmental Impact" of any FGD system is highly site specific.

1.4.4.6.1 Stack Emissions - Rainout from the plume did not occur in the CT-101 process (even during operation without reheat), principally because (1) the test stack was well insulated, (2) the design stack velocity was about 25 fps (too low to strip liquid from the stack and duct walls) and (3) there were two chevron mist eliminators and several bends in the flue gas duct downstream of the absorber (three 90° bends and one 180° bend).

1.4.4.6.2 Comparison with Direct Lime or Limestone Systems - Assuming that the CT-101 process has been designed to operate in a mode of water use minimization (water-savings) similar to that tested in September of 1976 at Scholz and that it includes a catalyst recovery waste disposal system like those installed on large systems in Japan, its secondary environmental impact, with respect to water quality, will be similar to that of direct lime or limestone systems if the site specific conditions are the same**. A possible exception is the concentration of total dissolved solids in the leachate from the waste area. The Chiyoda waste material is gypsum, which is approximately 50 times more soluble than the calcium sulfite produced in direct lime or limestone processes. On the other hand, if a magnesium-promoted direct lime or limestone system is compared with a CT-101 process, the total dissolved solids in the liquor included with the waste solids will be greater for the lime or limestone system since magnesium is significantly more soluble than calcium.

1.4.4.6.3 Waste Solids - The CT-101 waste solids, which were almost entirely gypsum or calcium sulfate ($CaSO_4 \cdot 2H_2O$), ranged in particle size from 0.02 to 0.10 mm and were typically dewatered by centrifuging to 85 wt% solids.

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* The exact quantitative effect on sulfur dioxide removal is not known, due to changing process conditions and the difficulty in correlating sulfur dioxide removal with several variables in such a large plant. However, Chiyoda's pilot plant experience indicates that on going from no magnesium in solution to 10,000 ppm in solution, at about 1500 ppmv inlet sulfur dioxide, removal efficiency will drop about 5%.

** Under many conditions there will be a free liquid blowdown from the prescrubber; however, several disposal or treatment options are available and it should be possible to handle this stream through proper design (See Sections 1.4.4.1.1 and 1.4.4.6.4C).

During the test program the solids content ranged from 82 to 88 wt%. The drier waste solids were produced toward the end of the program due to (1) reduction of the mother liquor acid concentration, (2) lowering the average limestone particle size, and (3) more stable process operation. The waste had the appearance of a moist powder or moist, fine sand and appeared somewhat drier than the CEA/ADL waste. The CT-101 waste solids content was typically high enough to enable landfill disposal and compaction as indicated by laboratory testing. Compaction of the waste solids in the laboratory was accomplished at the solids content routinely obtained by centrifuging.

A. Disposal Methods - The method used in disposing of the CT-101 solid waste at Scholz, which was identical to the CEA/ADL waste disposal method, was employed for expediency (See Section 1.3.4.6.3A). A photograph of the CT-101 solid waste pond is shown in Figure 1.4-8, and pond construction details are shown in Figure 1.3-10. At a full-scale installation of the CT-101 process it is likley that landfill, conventional ponding, or stacking (a type of ponding) would be employed.

(1) Landfill - Laboratory testing indicated that the CT-101 waste, as dewatered by centrifuging, could be compacted to produce a landfill with high strength and low short-term settlement comparable to a well compacted sandy clay. As with the CEA/ADL waste, long-term settlement (indicated by laboratory testing) may pose constraints on future structural use of the land-filled area. Laboratory testing also indicated that dry densities in excess of 90 pounds per cubic foot could be obtained by compaction in a landfill.

(2) Ponding - If the CT-101 waste is slurried to a pond, laboratory testing indicates that the dry density of the settled waste will be somewhat less than 70 pounds per cubic foot. This dry density indicates that CT-101 waste settled in a pond would occupy about 30% more volume than the same waste compacted in a landfill.

(3) Stacking - Stacking is a gypsum disposal method employed extensively by the phosphate industry and should be applicable to CT-101 waste. In the stacking operation an earthen starter dike is constructed to form a pond in which the gypsum slurry is deposited. Once sufficient gypsum is deposited in the pond, the gypsum is dug out with a dragline and placed around the perimeter of the pond to form subsequent starter dikes. Gypsum stacks over 100 feet tall with side slopes steeper than 1.5 horizontal to 1.0 vertical are common. No testing (other than comparing the soil mechanics properties of CT-101 waste with properties of phospho-gypsum) has been conducted to determine the feasibility of stacking CT-101 waste. Laboratory and field testing are planned for determining the feasibility of stacking CT-121 waste in conjunction with the CT-121 evaluation program that will be conducted at Scholz beginning in mid-1978.

B. Utilization Potential - Laboratory testing indicated that CT-101 waste could be used successfully as a soil amendment, for wallboard manufacture, and as a cement retarder. Agriculture utilization of CT-101 waste was demonstrated by selling essentially all of the waste that was produced during the last nine months of CT-101 process operation for use as a soil amendment. The waste was successfully spread and utilized as a calcium

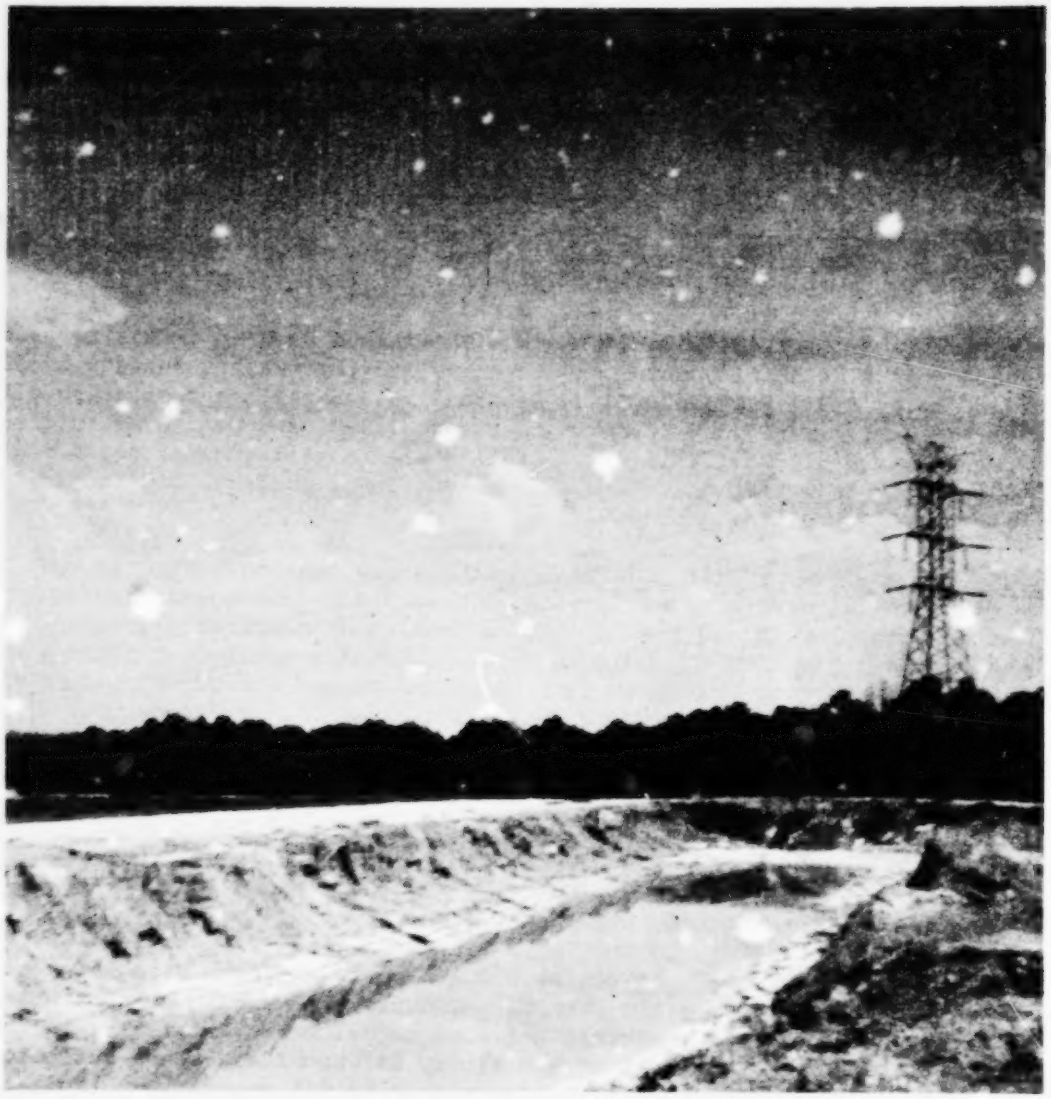

FIG. 1.4-8. CT-101 Dilute Acid Process Waste Pond

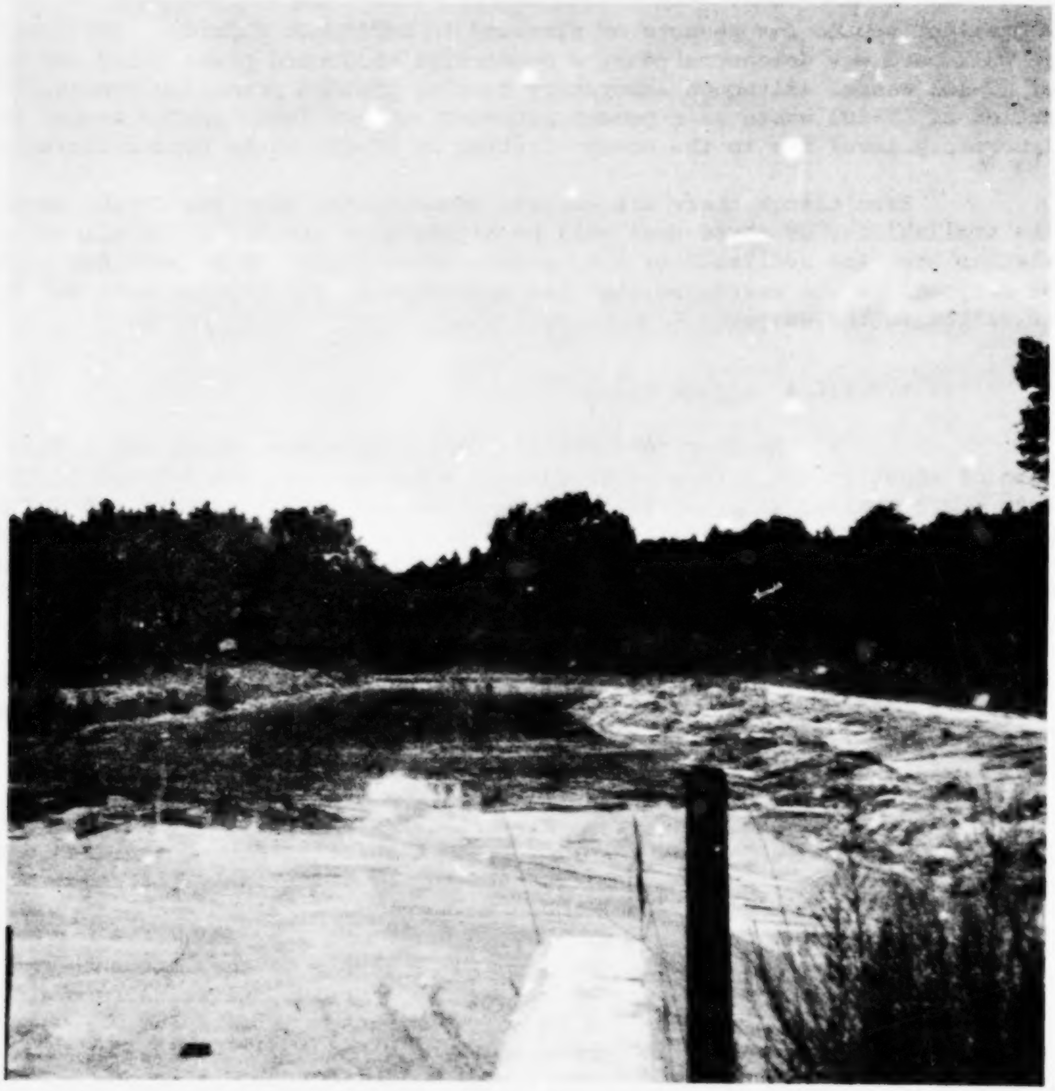

FIG. 1.4-7. CT-101 Dilute Acid Process Liquid Waste Settling Pond

and sulfur source for peanuts on farmland in northwest Florida. Utilization as wallboard was demonstrated at a commercial wallboard plant using 200 tons of CT-101 waste. Although laboratory testing yielded promising results, utilization of CT-101 waste as a cement retarder was not investigated beyond the laboratory level due to the short duration of CT-101 waste production at Scholz.

Even though there are several demonstrated uses for CT-101 waste, the availability of these uses will be highly site specific. At all sites, whether uses are available or not, provision will have to be made for storage or disposal of the waste, so that the operation of the process does not depend on marketing the waste.

1.4.4.6.4 Liquid Waste

A. Monitoring - The CT-101 liquid waste, which had a flow rate of about 20 gpm before water-savings modifications and 3-5 gpm (0.15-0.25 gpm/MW) after water-savings, was neutralized with limestone and passed through a settling pond (See Figure 1.4-7). The liquid waste settling pond overflow, which had been diluted with limestone slurry water which increased its flow rate to about 50 gpm, was monitored monthly for trace elements and major constituents. The monitoring parameters are listed in Table 1.3-7. Concentrations of total suspended solids in several samples of the CT-101 liquid waste settling pond overflow were in excess of the EPA effluent limitations for the low volume waste category of Steam Electric Power Generating Point Sources[4]. One or more samples had concentrations of boron, iron, manganese, selenium, chloride, nitrate, and sulfate in excess of the EPA Criteria for Water Supply[5,6]. The fluoride concentration of all monthly samples was in excess of U.S. Public Health Service Drinking Water Standards[7]. It is important to note that except for one atypical startup condition there were no violations of water quality standards or criteria in the receiving waters (where they are applicable) due to the CT-101 liquid waste at Scholz.

B. Source of Contaminants - Analysis of the inputs to the process indicated that essentially all of the fluoride, nitrogen, and selenium and about 95% of the chloride and mercury (which was detectable in several liquid waste samples) entering the process originated in the coal.

C. Treatment - A portion of the liquid waste settling pond overflow was concentrated by a factor of 100 using a pilot Resources Conservation Company (RCC) vapor compression evaporator (See Figure 1.4-9). Over 99% of the liquid waste treated was recovered as high quality water (less than 10 ppm total dissolved solids), and the remaining 1% (concentrate) had to be disposed. Reportedly, the concentrate (waste brine) can be mixed with dry flyash and lime for disposal in a landfill, but this has not been conclusively proven (laboratory testing by IU Conversion Systems is in progress). A conceptual design and economic study conducted by RCC[18] based on the pilot evaporator operation at Scholz, indicates that the capital and operating costs of concentrating the CT-101 waste using an RCC evaporator would be \$5.11 to \$8.71 per gallon/day of capacity and \$2.45 to \$3.59 per 1000 gallons treated, respectively. Brine disposal, for which an acceptable method will have to be found for a non-arid site, is not included in the above costs.

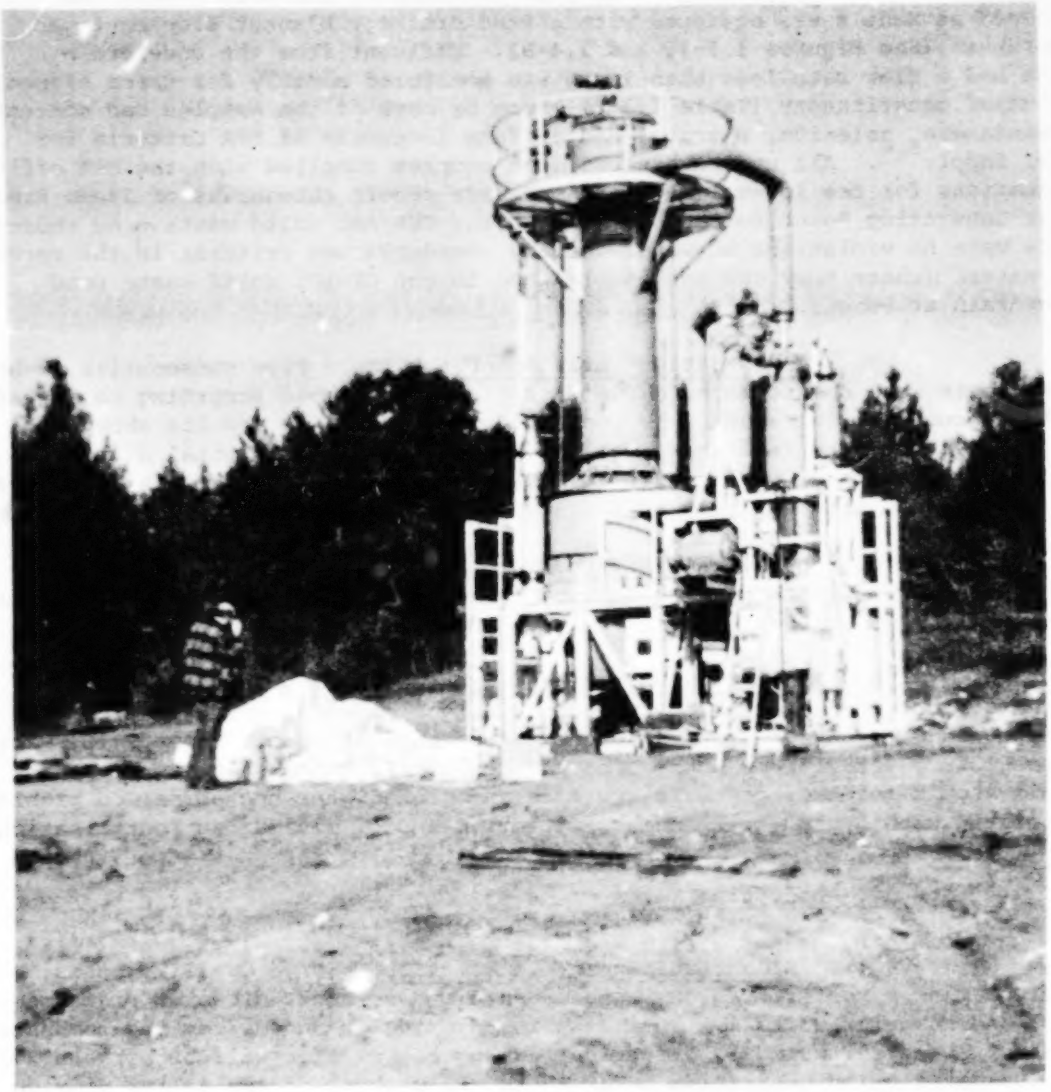

FIG. 1.4-9. Resources Conservation Company Pilot Evaporator Used To Concentrate CT-101 Liquid Waste at Scholz

1.4.4.6.5 Leachate and Runoff

A. Monitoring - The pond in which CT-101 solid waste was disposed at Scholz was equipped with a sand drainage blanket sloping to an underdrain (See Figures 1.3-10 and 1.4-8). Effluent from the underdrain which had a flow rate less than 1 gpm was monitored monthly for trace elements and major constituents (Table 1.3-7). One or more of the samples had concentrations of manganese, selenium, nitrate, and sulfate in excess of EPA Criteria for Water Supply[5,6]. All underdrain effluent samples complied with the EPA effluent limitations for the low-volume waste and area runoff categories of Steam Electric Power Generating Point Sources[4]. As with the CEA/ADL solid waste pond underdrain, there were no violations of water quality standards and criteria in the receiving waters (where they are applicable) due to the CT-101 solid waste pond underdrain at Scholz.

B. Concentration Variation With Time - Five consecutive 48-hour shake tests were conducted on a CT-101 solid waste sample according to a standardized procedure followed by IU Conversion Systems. The results shown in Table 1.4-6 indicate that there will be little reduction in total dissolved solids of leachate or runoff with time. The total dissolved solids concentration is controlled almost entirely by the solubility of gypsum.

C. Source of Contaminants - The analysis described above for the CT-101 liquid waste is also applicable to the CT-101 solid waste underdrain effluent.

D. Minimizing Surface and Groundwater Contaminants - The comments given in Section 1.3.4.6.4 concerning minimizing contamination of surface and groundwater are applicable to leachate or runoff from CT-101 waste disposal in a landfill or pond. The planned CT-121 waste stacking test at Scholz should identify potential surface and groundwater contamination problems associated with gypsum stacking and various means of minimizing contamination.

1.4.4.7 Operating Costs

1.4.4.7.1 Electrical Energy Consumption - For a CT-101 process incorporating a venturi, consuming between 4 and 5 inches of water pressure drop* and removing 85 to 90% of the sulfur dioxide produced from the combustion of about 3 wt% sulfur coal, electrical energy consumption will be between 3.2 and 3.5% of station generation. For 97 to 98% removal of the sulfur dioxide produced under the same conditions, electrical energy consumption will be about 4.0% of station generation. This assumes operation on a boiler producing about 2100 scfm of flue gas per megawatt of electrical generation and does not include ancillary support equipment such as limestone grinding and unloading or waste disposal. However, these should add not more than 0.1 to 0.2% to the total.

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* Only about 4 to 5 inches of water pressure drop are necessary in a venturi to saturate flue gas. For a system depending on a venturi for particulate removal, about 12 inches of pressure drop in the venturi might be typical. For this situation, the electrical energy consumption would increase about 0.5% over the above figures.

TABLE 1.4-6

CT-101 DILUTE ACID PROCESS-
ANALYSIS FROM FIVE CONSECUTIVE 48-HOUR SHAKE
TESTS CONDUCTED ON THE CT-101 WASTE SAMPLE AS PRODUCED

	WASH WATER CONCENTRATION (mg/l EXCEPT pH)				
PARAMETER	FIRST WASHING	SECOND WASHING	THIRD WASHING	FOURTH WASHING	FIFTH WASHING
pH	7.2	7.4	7.3	7.2	7.3
Phenolphthalein Alkalinity as $CaCO_3$	0	0	0	0	0
MO Alkalinity (Total) as $CaCO_3$	10	30	10	20	10
Hardness as $CaCO_3$	1660	1610	1540	1530	1520
Sulfite as SO_3	3	3	3	3	3
Sulfate as SO_4	1554	1526	1492	1394	1429
Chloride as Cl	6	1	1	1	1
Total Dissolved Solids	2634	2420	2340	2266	2228
Total Suspended Solids	62	172	1017	30	138

1.4.4.7.2 Reheat - The comments under Section 1.3.4.7.2, "Reheat," in the CEA/ADL discussion apply to the CT-101 process as well.

1.4.4.7.3 Limestone Utilization - Utilization of the calcium in the limestone averaged between 96 and 97% over the entire test program and was unaffected by normal process variations or operation on flue gas containing high levels of particulates. The actual limestone stoichiometric ratio for a full-scale process, including a catalyst recovery system and employing a water-savings mode of operation, would be on the order of approximately 1.05, versus the 1.04 indicated by the above utilization, since limestone would be required to neutralize the sulfate ion introduced with the catalyst makeup.

1.4.4.7.4 Catalyst Consumption - For operation of the CT-101 process on flue gas produced from the combustion of coal containing about 0.1 to 0.2 wt% chlorides and 2 to 3 wt% sulfur, catalyst consumption on a process including a catalyst recovery system (not used at Scholz, see Section 1.4.1) will be less than 0.2 mole % (1.2 wt%), of the sulfur dioxide removed as moles of $Fe_2(SO_4)_3$/mole of sulfur dioxide removed.

1.4.4.7.5 Other Operating Costs - The comments under Section 1.3.4.7.5 of the CEA/ADL discussion apply to the CT-101 process as well.

1.5 FW/BF Dry Adsorption Process

1.5.1 System Design - In the FW/BF process, Figure 1.5-1, flue gas passes through the adsorber, Figures 1.5-2 through -4. There it contacts parallel beds of char where sulfur dioxide, sulfur trioxide, particulates, and possibly some nitrogen oxides, are removed. The sulfur dioxide combines with water vapor and oxygen (also absorbed from the flue gas) and forms sulfuric acid. After becoming saturated with sulfuric acid, the char is circulated to a regeneration section, Figure 1.5-5. There, hot sand from a fluidized-bed sand heater (at Scholz, No. 2 fuel oil fired) is added to regenerate the sulfuric acid to sulfur dioxide. The regenerated char and sand are separated, the sand is reheated, the char is cooled, and the char is recirculated to the top of the adsorber for reuse. At Scholz, the exhaust gas from the sand heater is ducted to the boiler air preheater flue gas inlet for heat recovery. This also provides for the injection of the sulfur dioxide produced as a result of the combustion in the sand heater into the main flue gas stream entering the adsorber.

The off-gas stream from the regenerator (20 to 30% sulfur dioxide) is passed to the RESOX* section, Figure 1.5-6, where it is reacted with a bed of coal to produce elemental sulfur in the gaseous state. The sulfur is then condensed (in a shell-and-tube heat exchanger at Scholz, Figure 1.5-7), and stored for sale. The ash products produced in the RESOX vessel can be burned in the plant boiler to recover the high residual carbon values. The tail gases leaving the sulfur condenser consist of carbon dioxide, water, nitrogen and those remaining sulfur values not converted to elemental sulfur. At Scholz, these gases are recycled to the boiler (via a centrifugal blower) where the sulfur values are oxidized to sulfur dioxide. Future designs would probably include recycle of these gases to the sand heater combustion zone. However, the requirement for oxidizing the residual sulfur values to sulfur dioxide was not recognized soon enough to incorporate this into the design at Scholz.

The adsorption section also includes a liquid nitrogen tank/evaporator and distribution system. This system was added to the Scholz design at the recommendation of Bergbau. Its original purpose was to provide an inert gas blanket in the adsorber during shutdown operations. However, as indicated in Section 1.5.4.7.6, problems with temperature excursions in the char required much more extensive use of the nitrogen purge system at Scholz.

1.5.2 Process Chemistry - In the adsorption section of the FW/BF process, boiler flue gases containing nitrogen oxides, sulfur oxides, oxygen, water vapor, and particulate matter come into contact with activated carbon pellets (char) which are the adsorption media. This char is produced from air-oxidized bituminous coal by forming in an extrusion press, carbonizing at low temperatures, and activating with water vapor. Sulfur dioxide, oxygen, and water vapor are adsorbed onto the char, according to the following overall adsorption equation:

$$SO_2 + \tfrac{1}{2}O_2 + H_2O \longrightarrow H_2SO_4 \qquad (1)$$

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* "RESOX" is a registered trademark of the Foster Wheeler Energy Corporation.

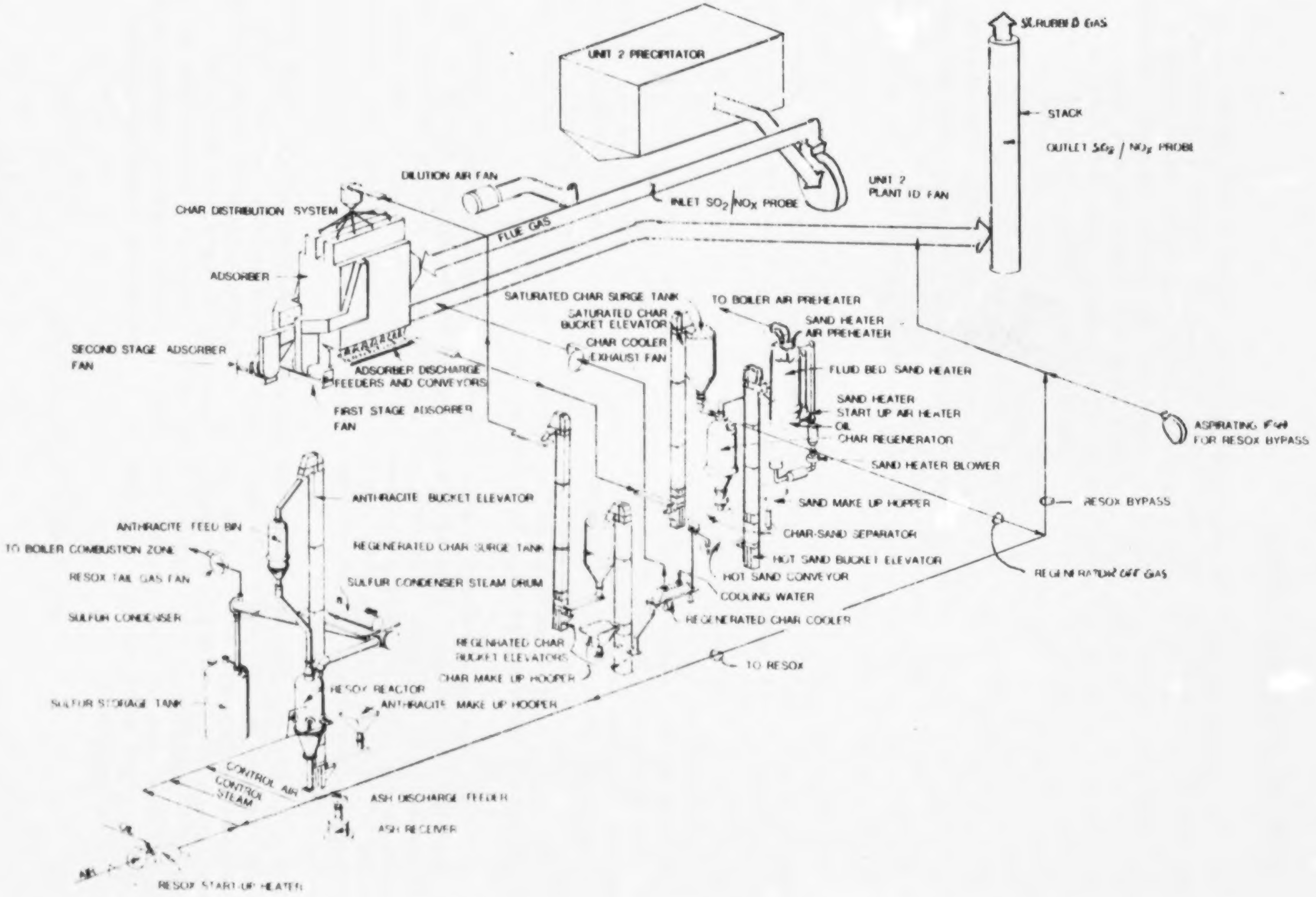

FIG. 1.5-1. FW/BF Dry Adsorption Process Flow Diagram

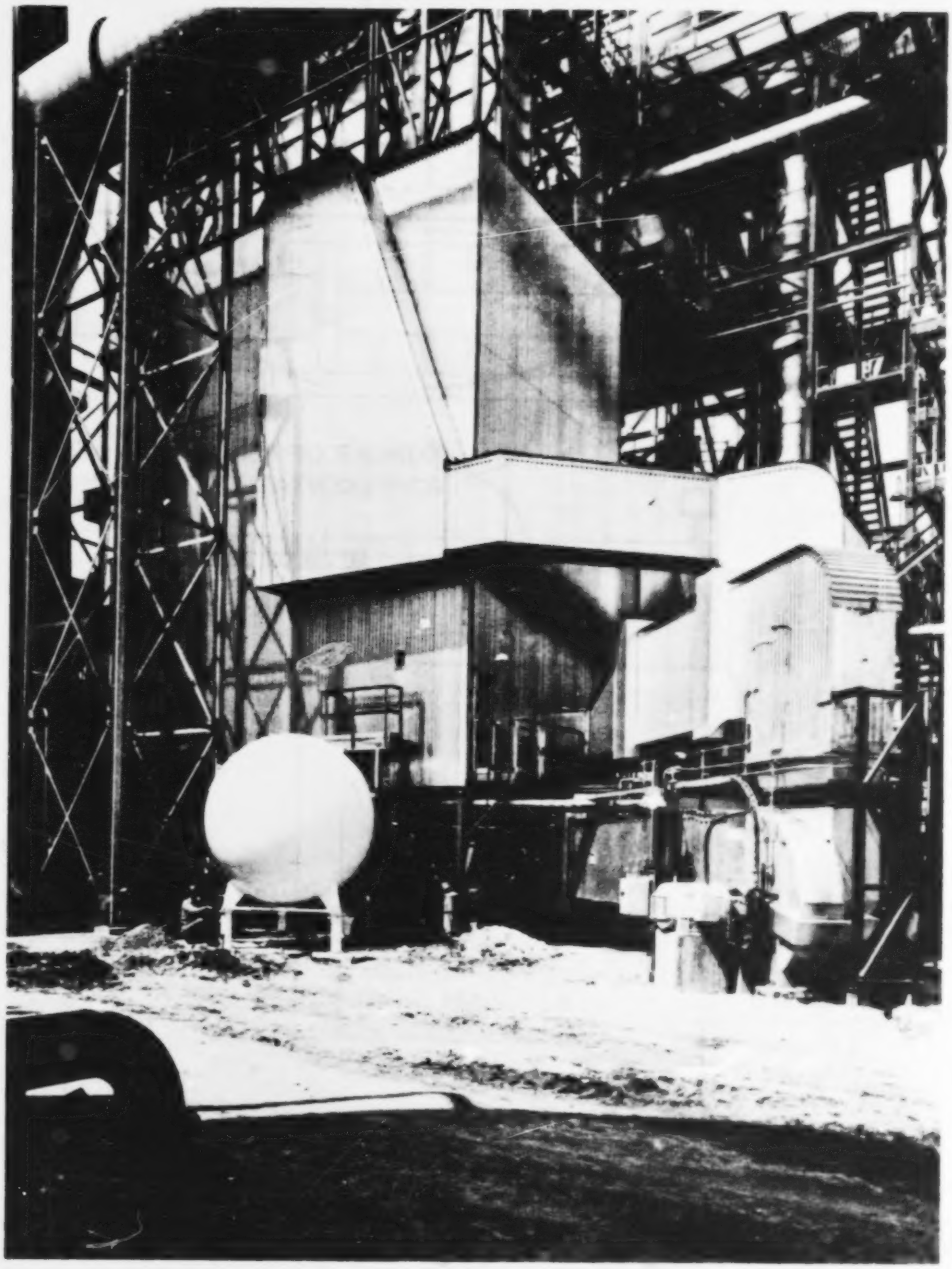

FIG. 1.5-2. FW/BF Dry Adsorption Process
Adsorption Section

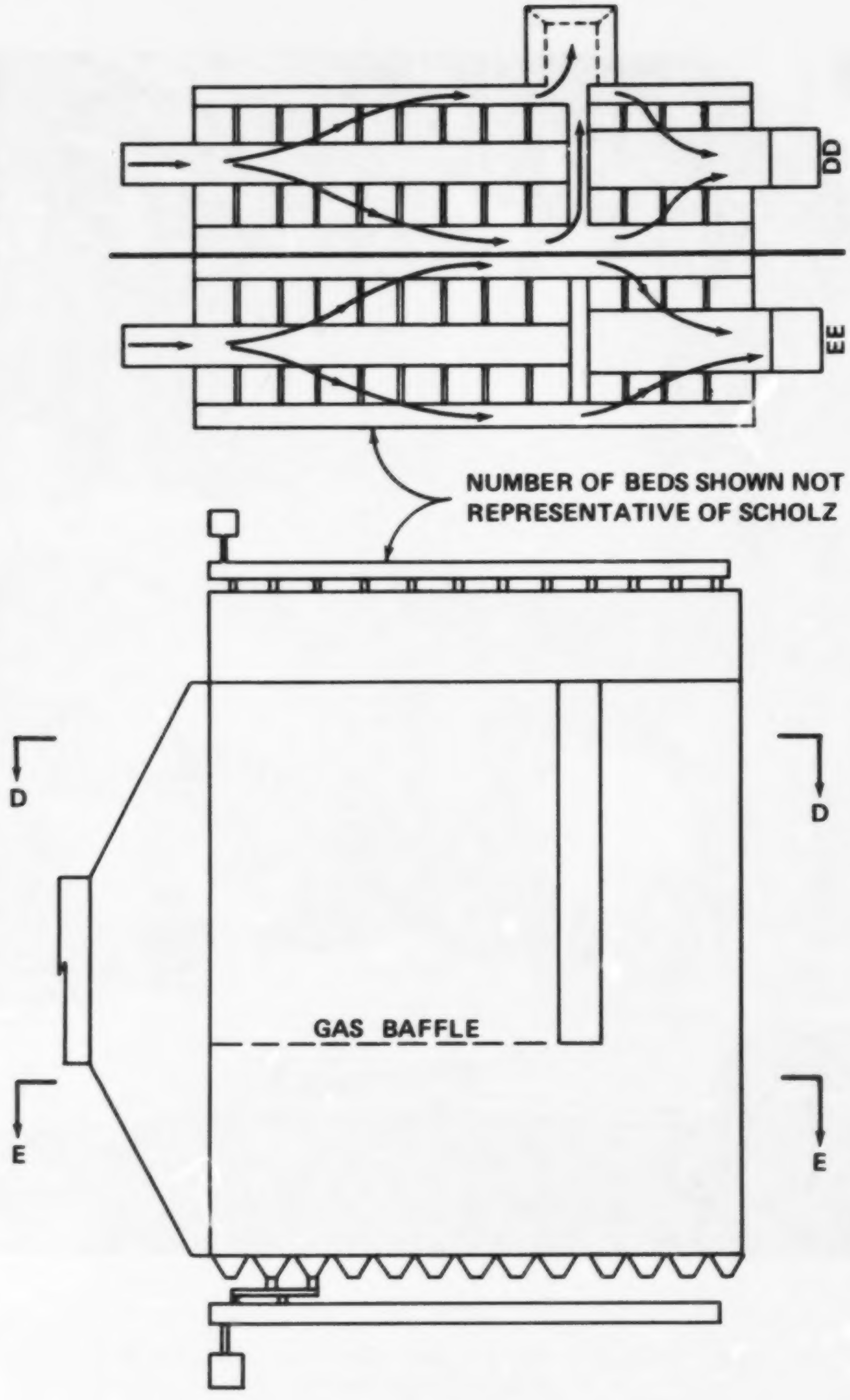

FIG. 1.5-3. FW/BF Dry Adsorption Process Typical Adsorber

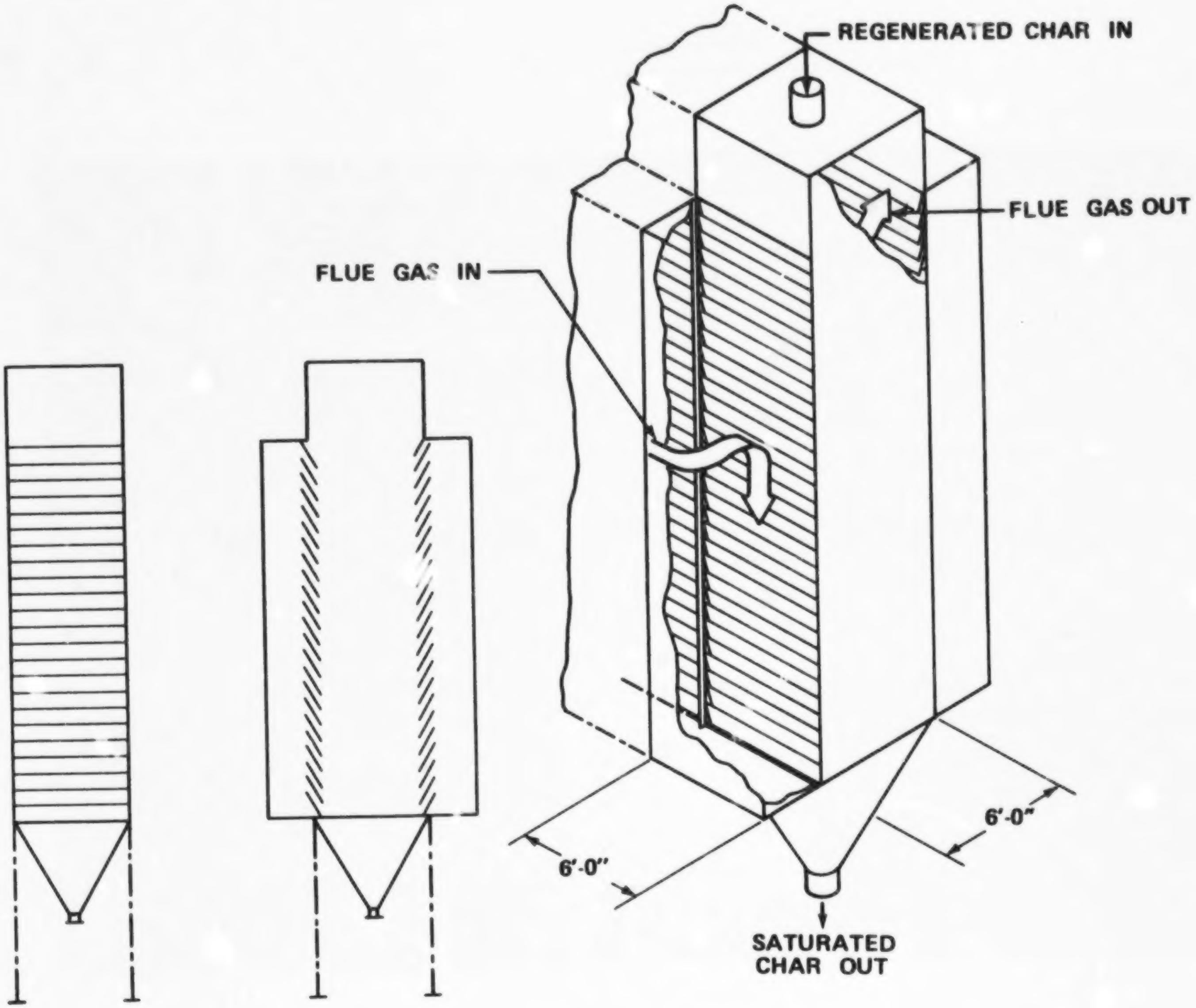

FIG. 1.5-4. FW/BF Dry Adsorption Process Absorber Module Detail

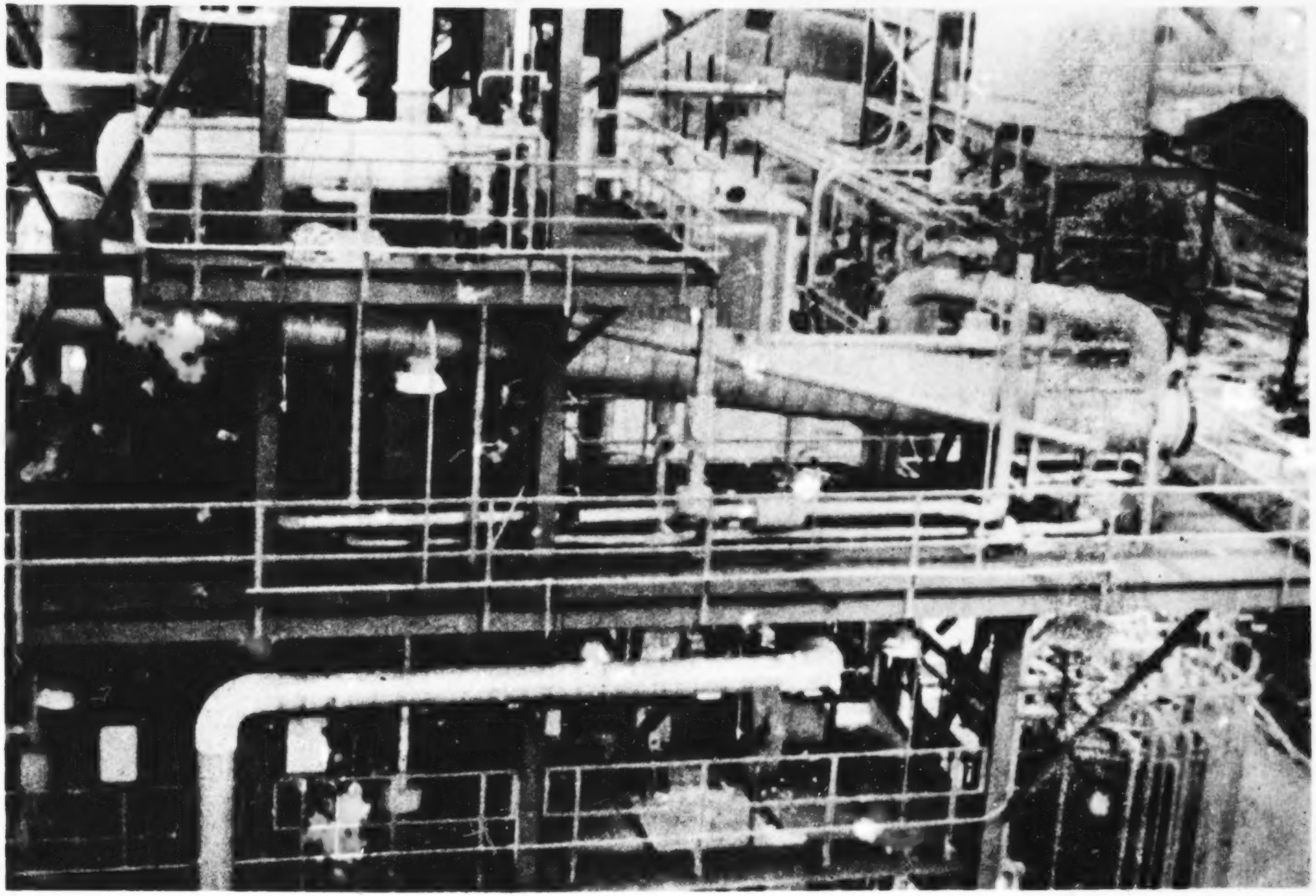

FIG. 1.5-7. FW/BF Dry Adsorption Process
Sulfur Condenser

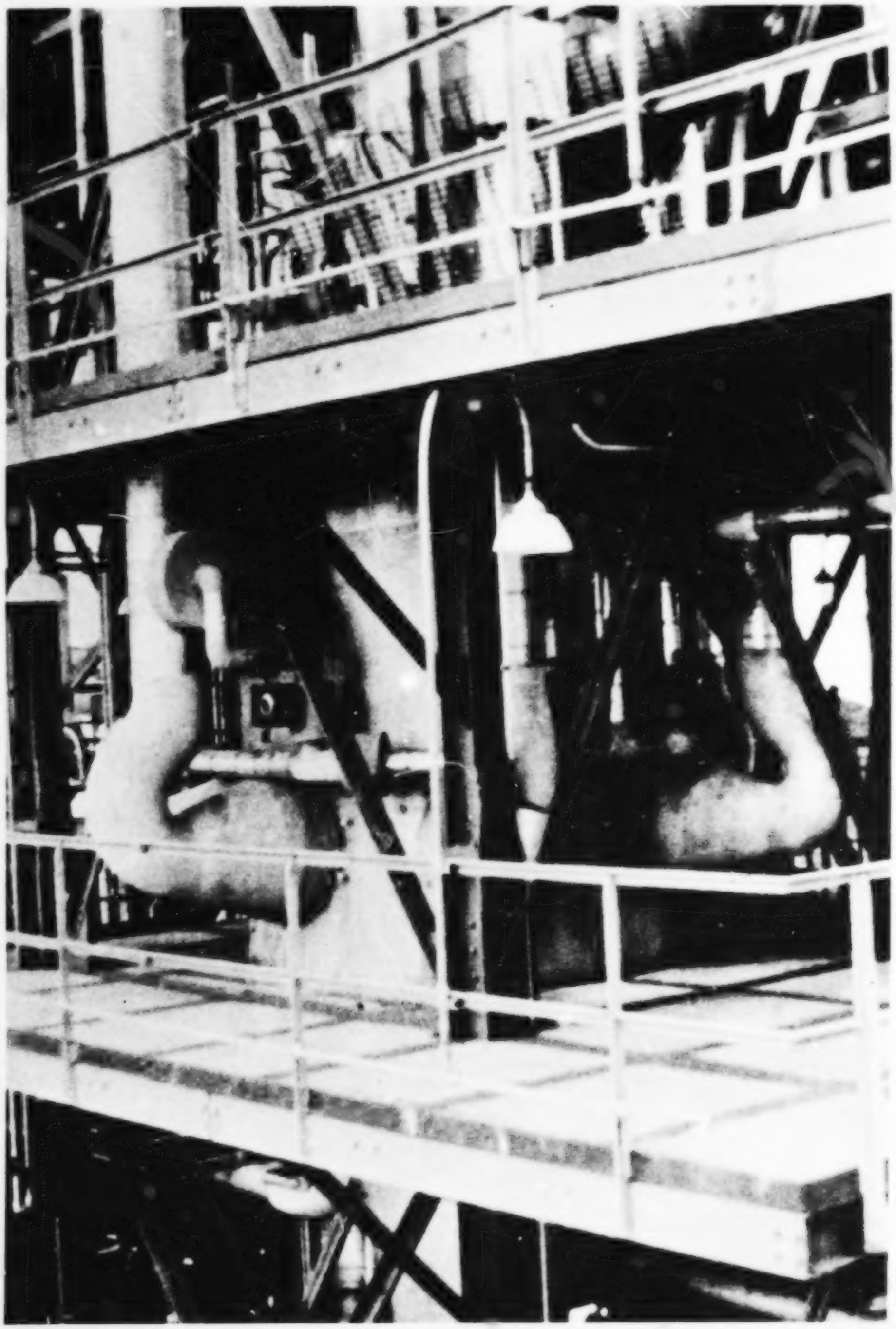

FIG. 1.5-6. FW/BF Dry Adsorption Process Resox Reactor

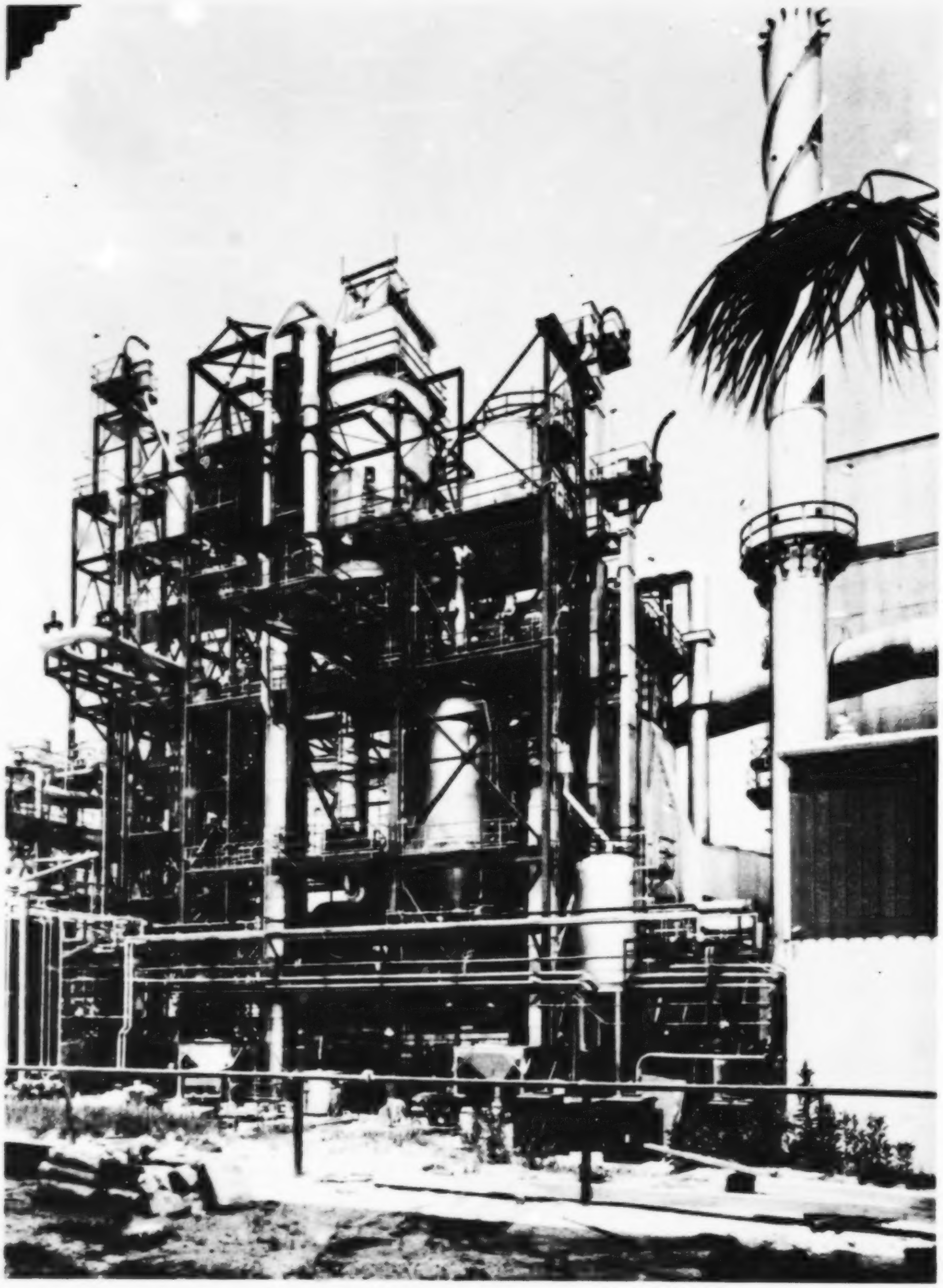

FIG. 1.5-5. FW/BF Dry Adsorption Process
             Regeneration Section

A detailed mechanistic discussion of this adsorption reaction is given in References 19 and 20. In addition, although no measurements were made at Scholz, sulfur trioxide is similarly adsorbed. Nitrogen oxides are supposedly adsorbed by a similar mechanism. As will be discussed later, limited data that seem to confirm this contention were collected at Scholz. However, the principles of the adsorption of nitrogen oxides are still under investigation. Particulate matter is collected on the surface of the char pellets which act as impingement filters due to their size and physical arrangement.

The char eventually becomes saturated with sulfuric acid and must be regenerated to recover its adsorptive capacity. This regeneration is accomplished by heating the char in an inert atmosphere to temperatures in excess of 1200°F. Under these conditions, the sulfuric acid is regenerated according to the following overall equation:

$$H_2SO_4 + \tfrac{1}{2}C \longrightarrow \tfrac{1}{2}CO_2 + H_2O + SO_2 \qquad (2)$$

Any nitrogen oxides that were adsorbed are thought to be reduced according to the following overall reaction:

$$2NO_x + C \longrightarrow xCO_2 + N_2 \qquad (3)$$

These reactions result in the production of a concentrated stream of off-gases, consisting of sulfur dioxide, carbon dioxide, water, and nitrogen, with a sulfur dioxide concentration in the range of 20 to 30% by weight.

The concentrated off-gases produced in the regeneration section are introduced in counterflow to a bed of crushed anthracite coal. The reactions occurring in this RESOX reactor are represented by the following overall equation:

$$SO_2 + C \longrightarrow CO_2 + S \qquad (4)$$

The sulfur is produced in the form of a gas which is subsequently condensed. The nitrogen and carbon dioxide constituents of the regenerator off-gas pass through the RESOX reactor without taking part in the reactions. Reaction 4 is not totally selective to sulfur. Side reactions can produce, among others, hydrogen sulfide. Therefore, the tail gases from the sulfur condenser must be burned to reduce residual sulfur values to sulfur dioxide. This residual sulfur dioxide can then be recycled to the adsorber inlet for removal. For a more complete discussion of the phenomena of nonselectivity and the importance of the various operating variables, refer to Reference 21.

1.5.3 System Operation - The operating time of the system is summarized in Table 1.5-1. The initial operation began in August, 1975 when flue gas was passed for a period of ten days, August 11 through 21. It quickly became obvious that the system was not ready for operation and that several design modifications were necessary before long-term operation could be considered. Several problem areas were identified, particularly difficulties with controlling heat buildup in the adsorber beds, and an extensive modifications/maintenance program was undertaken from October, 1975 to March, 1976.

TABLE 1.5-1

FW/BF DRY ADSORPTION PROCESS
OPERATING HISTORY

	OPERATING PERIOD (DAYS)			
	8/75	3-4/76	4-5/76	TOTALS
ADSORPTION	10	24	6	40
REGENERATION	6	23	6	35
SULFUR PRODUCTION	0	5	0	5
	TOTAL ELAPSED TIME			267

Flue gas was again introduced to the system on March 3, 1976, and operation continued for about three days. However, shutdown was again necessitated because of hot spots within the adsorber. After modifications in the operational procedures, the unit was restarted on March 15 and ran continuously through April 4, when it was shut down in anticipation of a boiler outage. During this period, the inlet sulfur dioxide concentration from the boiler ranged between 700 and 3200 ppmv (typical values ranged between 2000 and 2400 ppmv). The sulfur dioxide in the inlet gas was cooled to about 260°F and diluted to between 1700 and 2000 ppmv via the addition of air for cooling purposes. (Lower inlet temperatures improve sulfur dioxide adsorption and char-bed temperature control.) Outlet sulfur dioxide values ranged between 200 and 500 ppmv, giving typical sulfur dioxide removals in the range of 70 to 90%.

Operation resumed on April 27 and continued through final shutdown on May 3, 1976. The process was shut down due to a resumption of the problems with hot spots in the char beds (caused during an attempt to balance the booster fans while the system was on line) and the fact that the char consumption was running about five times anticipated levels. Sulfur dioxide removal data during this period are spotty, due to a malfunction of the outlet analyzer. However, spot checks of the sulfur dioxide concentrations in the gas show removal efficiencies generally in agreement with those of the March/April period.

1.5.4 System Performance

1.5.4.1 Reliability/Operability Potential - Neither the mechanical reliability nor the process operability was acceptable. Operation of the FW/BF system at Scholz was hampered throughout the test program by debilitating mechanical problems. The major ones were associated with an inability to control temperature buildup in the char in the adsorption section, the mechanical reliability of the char/sand separator, and operation of the RESOX system, particularly with respect to plugging of the sulfur condenser with a mixture of sulfur and carbon particles.

Bergbau has had better success with the adsorption and regeneration sections of the process at Lunen*. However, the system at Lunen includes a modified Claus unit for the production of sulfur from sulfur dioxide. RESOX, FW's sulfur dioxide reducing step, could not be successfully operated at Scholz, primarily because of the plugging mentioned above. Because of the unique potential of the RESOX technology, (It utilizes coal directly as a sulfur dioxide reductant) the EPRI is funding further development of the RESOX subsection on the Bergbau system at Lunen.

1.5.4.2 Sulfur Dioxide Removal Capability - As with most FGD systems, sulfur dioxide removal is simply a matter of designing sufficient capability into the system. For the short periods that the system operated, sulfur dioxide removal typically ranged between 70 and 90%.

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* Bergbau has successfully operated a 35 megawatt version of the process at the Kellerman Power Station of STEAG, AG near Lunen, West Germany, since February, 1975.

During the March/April period inlet sulfur dioxide concentrations from the boiler ranged between 700 and 3200 ppmv with typical values ranging between 2000 and 2400 ppmv. The sulfur dioxide in the inlet flue gas was diluted to between 1700 and 2000 ppmv and cooled to between 230 and 250°F via the addition of dilution air*. Outlet values ranged between 200 and 500 ppmv, giving sulfur dioxide removals of 70 to 90%. These sulfur dioxide removal efficiencies were achieved while treating combustion gases produced by burning approximately 3 wt% sulfur coal, circulating char through the system at approximately 10,000 lb/hr (about 58% of design) and treating between 50,000 and 75,000 acfm of flue gas (70 to 85% of design).

There were two levels of sulfur dioxide removal efficiencies for the system during this period. The first 80 to 90% removal occurred prior to a shutdown of the regeneration section on March 24 and the second 70 to 80% removal occurred subsequent to restart of the regeneration section on March 26. This higher sulfur dioxide breakthrough (or lower removal efficiency) was caused by the high sulfuric acid saturation levels produced in the char while it remained stationary. Sulfur dioxide removal began to increase toward the end of the operating period as the more saturated char was removed from the adsorber. In future installations, a designed over-capacity in the regeneration and RESOX sections may be desirable to minimize this recovery time. This decoupling of the adsorption and regeneration sections is a significant advantage for the process, since for short periods of time sulfur dioxide removal is not dependent on continuous operation in the regeneration and RESOX sections of the process.

1.5.4.3 Nitrogen Oxides Removal Capability - The nitrogen oxides (NO_x) removal capability of the FW/BF system has been the subject of some controversy. Bergbau's initial pilot plant tests at Welheim, West Germany, indicated nitrogen oxides removals on the order of 30 to 40%. However, operation of the Bergbau system at Lunen has not confirmed this. To date, no nitrogen oxides removal has been observed there.

During operation at Scholz, in the latter part of April and early May, 1976, nitric oxide (NO) removals were measured ranging between 17 and 50% and averaging approximately 20% of the 200 to 400 ppmv inlet nitric oxide concentration in the flue gas. It should be emphasized that strong claims for the validity of these data cannot be made, since verification by independent methods was not possible. However, the system at Scholz was operated at adsorption section temperatures approximately 30 to 40°F lower than those that are typical at Lunen (lower temperature tests are planned at Lunen). Lower temperatures are known to favor increased nitrogen oxide removal efficiency. Nevertheless, these are the first data collected since operation of the Welheim pilot plant that indicate some nitrogen oxides removal. Further work must be done to confirm its validity.

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* Ambient air was added to the flue gas, prior to its introduction to the adsorber, to control the temperature in the char beds.

1.5.4.4 Particulate Removal Capability and the Impact of Particulates on System Operation - Particulate removal on the order of 80% for inlet loadings of approximately 0.05 gr/dscf have been observed in the Bergbau system at Lunen. However, char abrasion in that system has been minimized through meticulous attention to detail in the design. In the system at Scholz, where char abrasion and consumption was on the order of five times that experienced at Lunen, the system is believed to have added particulate to the flue gas. Measurements at the system outlet during a simulation of normal operation* indicated outlet particulate loadings (which under microscopic examination appeared to be principally carbon) of about 0.05 gr/dscf. Because of the short duration of operations, no measurements were made on the inlet to the system and actual addition rates (if any) could not be determined. However, the flue gas discharged from the process was occasionally marginally visible.

1.5.4.5 Impact of Contaminants - Chloride and fluoride ions adsorbed from the flue gas have caused corrosion in the modified Claus unit at Lunen from condensation during shutdown periods. For this reason, a proprietary dry alkali absorbent has been installed in the off-gas line between regeneration and the Claus unit there. Similar problems with stress-related corrosion at welds in the 304 stainless steel off-gas line from regeneration to RESOX occurred at Scholz. This corrosion, too, probably resulted from acid condensation during shutdowns.

Operation was too erratic to observe any effects of trace contaminants on process performance. However, because the system is a dry adsorption type, it is unlikely that trace contaminants would have any significant effect on process performance. None have been noted at Lunen.

1.5.4.6 Secondary Environmental Impact - Although a problem at Scholz (See Section 1.5.4.4), particulate emissions have not been a problem at Lunen.

No experience was gained with the storage of elemental sulfur from the process at Scholz. However, any problems would be expected to be significantly less than those associated with disposal of waste from nonregenerable FGD systems.

1.5.4.7 Operating Costs

1.5.4.7.1 Electrical Energy Consumption - The electrical energy consumption for a properly designed FW/BF system operating on 2 to 4 wt% sulfur coal would be expected to be approximately 2.5% of station generation based on the results at Scholz. (This assumes operation on a boiler producing about 2100 scfm of flue gas per megawatt of electrical generation.) However, FW's estimates, based on a conceptual redesign of the system, utilizing the results at Lunen and Scholz, indicate an electrical energy consumption between 1.5 and 2.0% of station generation for future systems.

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* The simulation was necessary because the RESOX section was not operational. It was conducted by stopping operation of the regeneration section during the period of the test, but continuing to circulate char through the adsorption section by running it into surge tanks. The stoppage of operation of the regeneration section was necessary because to operate the regeneration section without RESOX requires venting of regenerator off-gas to the stack.

1.5.4.7.2 Heat Energy

A. Sand Heater Operation - Unlike wet FGD processes which may require reheating of exhaust gases, the FW/BF process does not require reheat. However, significant quantities of energy are required for replacing the heat lost in the sand heating system and in the regeneration of the char. At Scholz, the sand was heated in a No. 2 fuel oil fired, fluidized-bed sand heater. However, because of economic and supply constraints on oil, future systems will probably have to utilize coal for heating sand.

Data at Scholz were collected for only one set of conditions of char flow, gas flow, and sulfur value in the boiler fuel. These data indicate that for operation on 3.5 wt% sulfur coal with approximately 90% sulfur dioxide removal, energy consumption in the sand loop is approximately 4.3% of the heat input to the power station. This assumes a thermal efficiency of about 35% at the station.

The energy consumption in the regeneration loop is directly proportional to the amount of sulfur removed. In the range of 2 to 4 wt% sulfur in the fuel, energy consumption is a function of the fuel sulfur content (at about 90% sulfur dioxide removal) can be estimated via linear interpolation from the above data.

B. RESOX Startup Heater - During operation of the RESOX section at Scholz, the RESOX startup heater was fired to supplement heat generation in the reaction zone of the RESOX vessel at low-load conditions. This was necessary because the turndown of the RESOX section of the process is limited to about 50% of design. (The RESOX section was sized for the full 40 megawatts of the Unit 2 boiler.) This startup heater consumed an additional 10 to 15% of heat energy above that used in the sand heater package. However, in all probability, this additional energy usage can be eliminated, via the use of multiple modules to facilitate turndown, if further development of the RESOX section is successful (See Section 1.5.4.1).

C. Steam Usage - Steam consumption averaged 500 to 600 lb/hr. Of this, 300 to 400 lb/hr was consumed as an inerting gas in the RESOX startup heater exhaust. This usage was added in 1975 after initial operation indicated startup heater exhaust gases were too oxygen rich. Future designs would undoubtedly eliminate this requirement. The remaining 200 lb/hr was consumed as control steam in the RESOX vessel and for steam tracing. This 200 lb/hr is equivalent to approximately 0.1% of the station energy input. Future designs could reduce this requirement by using steam produced in the sulfur condenser.

1.5.4.7.3 Char Consumption - The FW/BF system at Scholz was originally anticipated to consume 0.12 pounds of char per pound of sulfur dioxide removed. The actual char consumption at Scholz was approximately 0.64 pounds of char per pound of sulfur dioxide removed, due to excessive mechanical attrition.

Experience to date at Lunen has shown char consumptions of approximately 0.18 pounds of char per pound of sulfur dioxide removed. However, since the earlier work at Lunen, Bergbau has begun testing a more abrasion resistant char which is expected to lower the consumption somewhat.

1.5.4.7.4 Sand Consumption - Sand consumption during periods of operation in March and April of 1976 averaged about 260 lb/hr. This contrasts with the 15 lb/hr figure (0.03% of the design sand circulation rate) anticipated for normal process losses by the sand heater manufacturer. It is impossible to determine how much of this sand consumption was due to normal process losses and how much was due to "unaccounted for losses" that can be corrected in the future. However, sand is one of the minor cost items for operation of the system.

1.5.4.7.5 Anthracite Consumption (RESOX) - The best estimate, from the sporadic data gathered at Scholz, of anthracite consumption is about 0.9 pounds of anthracite per pound of sulfur dioxide delivered to RESOX or approximately twice the design consumption rate. However, it is unrealistic to draw firm conclusions about the actual anthracite consumption in any future installations from these data, since the RESOX concept requires significant further development (See Section 1.5.4.1).

1.5.4.7.6 Nitrogen - At Scholz, nitrogen was used almost continuously while the process was operating and during shutdowns, not only as an interting gas, but also as a cooling medium to control hot spots in the adsorber beds. During the 500-hour operating period in March and April, 1976, the nitrogen consumption ranged between 0.75 and 1.0 tons per day. However, during shutdown periods, consumption ranged between 8 and 12 tons per day.

At Lunen nitrogen consumption has been considerably less. No nitrogen is required while the process is operating. During shutdowns a charge of about 1400 pounds is used to blanket the adsorber. After that, only about 10 pounds per day is required to maintain the blanket.

1.5.4.7.7 Other Operating Costs - From the operation of the system at Scholz, it is felt that mechanical maintenance requirements on the FW system will be on the order of 50% higher than those for either the dual alkali or CT-101 processes. However, operation at Lunen should be a better measure of these requirements. Operational manpower requirements will be similar to that of the other systems at Scholz.

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